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Pilot Scale Assessment of the Biological Fluidized Bed Process for Municipal Wastewater Treatment

Sewage Collection and Treatment
Report SCAT - 11

NHA 5484 81/12

85-6490/

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PILOT-SCALE ASSESSMENT OF THE BIOLOGICAL FLUIDIZED BED PROCESS FOR MUNICIPAL WASTEWATER TREATMENT

by

Dearborn Environmental Consulting Services
Mississauga, Ontario

for

Canada Mortgage and Housing Corporation

Report SCAT-11

This document may be obtained from -

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REVIEW NOTICE

This project was carried out under contract for the Policy Development and Research Sector, Canada Mortgage and Housing Corporation, through a grant provided under Part V of the National Housing Act. Technical guidance and assistance was provided by Mr. J.P. Stephenson, Environment Canada.

This report has been reviewed by Canada Mortgage and Housing Corporation and Environment Canada. The views, conclusions and recommendations expressed herein are those of the author and do not necessarily reflect the views and policies of either the Corporation or Environment Canada. Mention of trade names or commercial products does not constitute an endorsement or recommendation for use.

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Cat. No. En 43-51/11E
ISBN 0-662-11820-0
BEAUREGARD PRESS LIMITED

ABSTRACT

Pilot-scale investigations of the oxygenic biological fluidized bed process demonstrated that the system is a technically feasible alternative for carbon oxidation and nitrification of municipal wastewaters.

The design organic loading of the fluidized bed is limited by the ability of the process to remove suspended particulate matter. A treated effluent containing less than 15 mg/L filterable BOD_5 was produced consistently at a loading rate of 10 kg $BOD_5/m^3 \cdot d$, an order-of-magnitude higher than typical for conventional suspended growth systems. To attain less than 15 mg/L total BOD_5 content, without effluent polishing, required operation at a volumetric loading similar to conventional suspended growth systems, due to the contribution of suspended solids to the final effluent BOD_5. The fluidized bed responded rapidly to changes in influent conditions; the process was less susceptible to influent disturbances than the activated sludge and rotating biological contactor processes in terms of effluent filterable organic carbon concentration. There was significant deterioration in effluent quality in terms of suspended solids concentration under diurnally varying flow and concentration conditions.

Separate stage nitrification studies did not demonstrate the order-of-magnitude increases in volumetric ammonia oxidation rates claimed by other investigators, primarily because of the relatively low biomass concentrations possible during the treatment of high quality secondary effluent ($BOD/TKN < 0.4$). In addition, diffusional limitations in the reactor biofilms were identified, which significantly increased the reactor hydraulic retention time necessary for low effluent ammonia-nitrogen concentrations (≤ 1 mg/L).

Despite the apparent limitations associated with suspended solids removal during carbon oxidation and biomass concentration during separate stage nitrification of high quality secondary effluent, the process demonstrated a potential for application in such specific areas as:

1) expanding an existing suspended growth treatment facility;
2) upgrading an existing secondary treatment system to provide nitrification through fluidized bed pretreatment for organic carbon removal and/or to provide biological denitrification;
3) pre-treatment of high-strength industrial wastes.

It is recommended that further investigations of the process be concentrated in these areas.

RÉSUMÉ

On a démontré, à l'échelle pilote, la faisabilité des lits fluidisés anérobies pour l'oxydation du carbone et la nitrification des eaux résiduaires urbaines.

L'aptitude du lit fluidisé à éliminer les particules en suspension en limite la charge organique théorique. Un effluent traité contenant moins de 15 mg/L de DBO_5 filtrable a été obtenu uniformément pour une charge de 10 kg de DBO_5/m^3 et par jour, un ordre de grandeur supérieur à la normale pour les systèmes classiques de croissance en suspension. Pour obtenir une DBO_5 inférieure à 15 mg/L sans polissage de l'effluent, il a fallu que la charge soit semblable à celle des systèmes classiques, parce que les matières en suspension contribuaient à la DBO_5 de l'effluent final. Le lit fluidisé a réagi rapidement aux changements de la qualité des eaux à traiter, qui lui ont moins nui qu'au procédé aux boues activées et qu'aux lits de contact biologique rotatif, pour ce qui est de la concentration du carbone organique filtrable dans l'effluent. La qualité de l'effluent s'est gravement détériorée pour ce qui est de la concentration des matières en suspension, lorsque le débit et les concentrations variaient journalièrement.

Les études sur la nitrification à une étape distincte du traitement n'ont pas permis de confirmer les augmentations d'un ordre de grandeur de la vitesse d'oxydation de l'ammoniaque, revendiquées par d'autres chercheurs, principalement à cause des concentrations relativement faibles de la biomasse qu'il est possible de maintenir durant le traitement d'un effluent secondaire de grande qualité (DBO/azote total de Kjeldahl <0,4). De plus, on a déterminé les limites à la diffusion des pellicules biologiques des réacteurs, qui augmentent considérablement dans ces derniers le temps de rétention hydraulique nécessaire pour obtenir de faibles concentrations d'azote ammoniacal dans l'effluent (<1 mg/L).

Malgré les limitations apparentes qui sont associées à l'élimination des matières en suspension pendant l'oxydation du carbone et à la concentration de la biomasse pendant la nitrification séparée de l'effluent secondaire de grande qualité, le procédé pourrait servir dans des cas particuliers comme:

1) l'expansion d'une installation de traitement par croissance en suspension;
2) l'amélioration du traitement secondaire par ajout d'un dispositif de nitrification au moyen du prétraitement en lit fluidisé pour éliminer le carbone organique ou pour assurer la dénitrification biologique;

3) le prétraitement d'eaux résiduaires industrielles concentrées.

Il est recommandé que les études sur ce procédé se concentrent dorénavant sur ces cas.

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CONCLUSIONS

The results of the pilot-scale studies at the Wastewater Technology Centre in Burlington, Ontario, demonstrated the oxygenic fluidized bed process to be a technically feasible alternative to conventional biological treatment processes for carbon oxidation and nitrification of municipal wastewaters. The high removal rates attainable per unit volume could result in significant economic advantages in specific instances due to reduced reactor volumes along with the elimination of secondary clarifiers. However, the pilot-scale results also demonstrated significant limitations associated with fluidized bed technology.

The following conclusions regarding the biological fluidized bed process can be drawn:

Carbon Oxidation of Domestic Wastewater

1) The design loading of the fluidized bed process for carbon oxidation is limited by the ability of the system to effect suspended solids removal. Filterable effluent BOD_5 concentrations of less than 15 mg/L could be consistently attained at an organic loading of 10 kg $BOD_5/m^3 \cdot d$, an order of magnitude higher than typically specified for conventional suspended growth systems. However, under these loading conditions, the effluent contained an additional 20 mg/L total BOD_5 associated with the suspended solids. An effluent quality of less than 15 mg/L total BOD_5 without effluent polishing requires operation of the process at volumetric loadings similar to conventional suspended growth systems. A treated effluent containing less than 15 mg/L suspended solids could not be consistently produced at the Burlington pilot plant.
2) The suspended solids in the fluidized bed effluent represent, to a large extent, influent solids not removed in the biological reactor rather than biomass sheared from the support media. These suspended solids exhibited low settling velocities. Chemical coagulation improved the efficiency of gravity clarification; more work would be necessary, however, to assess the feasibility of this alternative. Dual media gravity filtration was found suitable for upgrading effluent quality to more stringent discharge requirements. The area associated with the filters represents three to six times the area associated with the fluidized bed reactor.
3) Operation of the process under diurnal loading conditions did not adversely affect the system's capability to effect carbon oxidation. Diurnal variations in waste

strength and flow did cause significant deterioration in effluent quality in terms of suspended solids and total BOD_5.

4) Although effluent quality remained relatively consistent in terms of the filtered organic carbon concentration (filtered BOD_5), the effluent suspended solids concentration varied greatly, depending on the influent concentration and process loading.

5) The high biomass concentrations reported by other investigators of fluidized bed processes were confirmed by the pilot plant studies. Volatile solids concentrations in the bed ranged from approximately 8 g/L to more than 15 g/L depending on the operating conditions. Other parameters describing fluidized bed operation, such as solids yield and oxygen utilization, are comparable to values cited for conventional systems operated at similar SRT.

6) No apparent temperature dependency was noted during carbon oxidation within the temperature range investigated.

7) The response time to changes in influent strength and loading is more rapid than conventional activated sludge or RBC processes. Based on the relative process gains, the latter processes are also more susceptible to influent disturbances than is the fluidized bed process in terms of the effluent filterable organic carbon concentration.

8) The fluidized bed process may not be applicable as a secondary treatment alternative to meet stringent receiving water discharge criteria ($\leq$15 mg/L BOD_5 and suspended solids) due to the problems in maintaining low suspended solids concentrations in the treated effluent. However, the high-rate nature of the process makes it attractive for such specific applications as:

    a) expanding an existing suspended growth secondary treatment facility to increase capacity for carbon oxidation;

    b) upgrading a secondary treatment system to provide nitrification through fluidized bed pretreatment for organic carbon removal and/or to provide biological denitrification;

    c) pre-treatment of high strength industrial wastes to allow discharge to municipal sewer systems.

Separate Stage Nitrification of Secondary Effluent

1) The separate stage nitrification studies did not verify the order-of-magnitude reductions in reactor volume requirements claimed by other investigators. Although

high specific nitrification rates were attained, the reactor biomass concentrations were considerably lower. The low organic carbon content of the secondary effluent feed to the Burlington system ($BOD_5/TKN < 0.4$) appears to be the principal cause of the lower biomass concentrations.

2) Significant diffusional limitations were identified within the fluidized bed biofilms. The resultant half-order ammonia-nitrogen concentration dependency, in conjunction with the CSTR mixing regime identified in the fluidized bed system under recycle conditions, significantly affects process design. The nominal hydraulic retention time required to attain an effluent ammonia-nitrogen concentration of 1 mg/L would be three to five times greater than that required to meet a discharge requirement of 5 mg/L NH_3-N, depending on the influent concentration.

3) Other parameters associated with biological nitrification, e.g., alkalinity consumption, oxygen utilization and solids yield, are comparable to values typically cited for similar suspended growth biological nitrification processes.

4) Based on a comparison of the Burlington studies and other separate stage fluidized bed nitrification investigations, there appears to be considerable advantage to operation of the process at a higher carbon:nitrogen ratio ($BOD_5/TKN \simeq 2$). The higher BVS concentration so attained more than compensates for the lower specific nitrification rates, and results in an overall reduction in the reactor volume required to produce a desired effluent quality.

RECOMMENDATIONS

Further investigations relating to the fluidized bed biological treatment process should be concentrated in the specific areas where the process has a demonstrated potential. These include:

1) large-scale demonstration of the technical and economic feasibility of applying fluidized bed technology to expand and/or upgrade an existing secondary treatment plant;

2) large-scale demonstration of the technical and economic feasibility of applying the fluidized bed process to denitrify municipal wastewaters from an existing secondary treatment facility; and

3) assessment of the feasibility of utilizing the biological fluidized bed for pretreatment of high strength industrial wastes prior to discharge to municipal sewer systems.

1 INTRODUCTION

The biological fluidized bed treatment process has been developed as a modification of more conventional fixed film biological systems (e.g., trickling filter, packed bed reactor or rotating biological contactor (RBC)) for carbon oxidation, nitrification and denitrification of municipal and industrial wastewaters. The basic concept involves passing a wastewater upward through a bed of such granular media as sand at a velocity sufficient to expand the bed beyond the point at which the frictional drag is equal to the net downward force of gravity. The upflow velocity produces a fluidized state of motion. A population of micro-organisms is developed as an attached film on the surface of the fluidized media.

Since the early investigative work of Beer et al(1,2), researchers have cited numerous potential advantages related to the fluidized bed concept, including:

1) higher reactor biomass concentrations than in conventional suspended growth processes,
2) lower pressure drops compared to packed bed systems, and elimination of the potential plugging problems in such systems,
3) possible elimination of secondary clarification for solids separation,
4) high mass transfer rates compared to other fixed film contactors due to the high exposed biomass surface area, and
5) reduced reactor volumes and land requirements for full-scale treatment plants.

To investigate these claims, Dearborn Environmental Consulting Services were retained by Canada Mortgage and Housing Corporation (CMHC) to co-ordinate a pilot-scale evaluation of fluidized bed technology. The project was conducted at the Environmental Protection Service's Wastewater Technology Centre (WTC) in Burlington, Ontario.

1.1 Study Objectives

Overall objective was to determine the key parameters governing the process design of the biological fluidized bed reactor for carbon oxidation and nitrification of municipal wastewater.

Specifically, the program was designed to:

1) characterize the process response for a range of hydraulic and substrate loading conditions for carbon oxidation and for nitrification,
2) characterize the process response to wastewater temperature,
3) assess the need for separation of suspended solids from the reactor effluent, and
4) assess the long-term variability of effluent quality from the system.

1.2 Experimental Program

To fulfill the desired objectives, a four-phase experimental program was developed:

Phase I:	'Steady-State' Response of the Biological Fluidized Bed Process in the Carbon Oxidation Mode.
Phase II:	Effect of Diurnal Loading Conditions on Process Performance in the Carbon Oxidation Mode.
Phase III:	Dynamic Response of the System in the Carbon Oxidation Mode.
Phase IV:	Process Response in the Nitrification Mode.

During these phases, miscellaneous factors affecting the process design, performance and economics were evaluated. These included characteristics of the process effluent suspended solids, oxygen utilization rate, and solids production rate.

Detailed discussions of the experimental program for each phase are in the relevant sections of the report.

2 OVERVIEW OF BIOLOGICAL FLUIDIZED BED TREATMENT TECHNOLOGY

Since the early 1970's when Beer and his co-workers(1,2) reported on the potential advantages of the biological fluidized bed process for denitrification, numerous researchers have become involved in evaluating the technology for treatment of potable water(3,4,5) as well as domestic(6,7,8,9,10) and industrial(11,12,13,14) wastewaters. All of the basic biological treatment processes -- carbon oxidation, nitrification and denitrification -- have been assessed during these investigations. The process has been evaluated in the anaerobic(15,16), anoxic(6,8,10), aerobic(17,18) (using air as the source of dissolved oxygen) and oxygenic(7,9,11) (using pure oxygen) operating modes.

Most fluidized bed research has concentrated on sand as the support medium. Activated carbon(8,19) and other granular media(15) have also been evaluated. In the United Kingdom, considerable development work has been done with porous support media in a three-phase flow mode(20), based on the fermentation research conducted by Atkinson and his co-workers(21,22).

This overview deals with the results of investigations of process performance in the oxygenic operating mode for carbon oxidation and/or nitrification. A more extensive review of the technical literature has been presented elsewhere by Cooper and Wheeldon(23).

Most of the oxygenic biological fluidized bed development work has been conducted by Dorr-Oliver Inc. and Ecolotrol Inc. The Water Research Centre in Stevenage, England is at present evaluating a two-stage fluidized bed system, including an oxygenic reactor for carbon oxidation and nitrification of domestic wastewater, and have published some preliminary performance data(10).

2.1 Carbon Oxidation in the Biological Fluidized Bed Process

Jeris et al(7,24) documented the high-rate treatment potential of the process for carbon oxidation of wastewaters based on extensive pilot-scale experimentation at Nassau County, New York. The pilot plant incorporated an "aeration cone" external to the fluidized bed reactor to dissolve a high concentration of pure oxygen in the raw wastewater. In the 0.95 m^3 reactor, volatile solids concentrations averaging 14 200 mg/L were attained using silica sand, with an effective size (d_{10}) of 0.5 mm and a uniformity coefficient (d_{60}/d_{10}) of 1.4, as the support medium. The average performance of the reactor for carbonaceous BOD_5 removal over 30 days is summarized in Table 2.1. Based

TABLE 2.1 SUMMARY OF FLUIDIZED BED TREATMENT PERFORMANCE FOR CARBONACEOUS BOD_5 REMOVAL*

Primary effluent flow (L/min)	59
Recycle flow (L/min)	131
Hydraulic retention time (min)	16
Temperature (°C)	24
Raw sewage BOD_5 (mg/L)	203
Primary effluent BOD_5 (mg/L)	94
Fluidized bed - total effluent BOD_5 (mg/L)	15
Fluidized bed - filtered effluent BOD_5 (mg/L)	4
% BOD_5 removal (raw to total effluent)	93
% BOD_5 removal (raw to filtered effluent)	99
Raw wastewater TSS (mg/L)	243
Primary effluent TSS (mg/L)	65
Fluidized bed effluent TSS (mg/L)	25
% TSS removal (raw to total effluent)	90

*Jeris et al(7)

on these operating conditions, the BOD_5 loading was approximately 0.6 kg/kg•d. The authors concluded that treatment performance equivalent to secondary effluent quality (BOD_5 <30 mg/L; suspended solids <30 mg/L) could be attained in a fluidized bed reactor providing approximately 16 minutes nominal retention time (based on the empty bed reactor volume) without effluent clarification.

Dorr-Oliver Inc. have extended the development work on oxygenic fluidized bed treatment through the application of their Oxitron system to municipal(9,25) and industrial wastes(11,26). Pilot-scale studies have been conducted at Orillia, Ontario and Greenwich, Connecticut on municipal wastewaters and at Cedar Rapids, Iowa on a concentrated carbohydrate wastewater from a corn wet-milling operation. These confirmed high carbon oxidation efficiencies associated with bed volatile solids concentrations in the range of 10 000 to 20 000 mg/L. Based on the Orillia operating data(25), the fluidized bed process produced an effluent containing 20 mg/L BOD_5 and 30 mg/L suspended solids at a loading of 3 kg unfiltered BOD_5/m^3•d, compared to a typical design value of 0.45 kg/m^3•d for conventional air activated sludge systems. It was also concluded, from comparison of Orillia and Nassau County data, that the design organic loading rate is a function of the influent characteristics(25). The BOD_5 treatment performance depended most on the process ability to remove suspended solids from the raw wastewater, as residual solids in the treated effluent contribute significantly to the final effluent total BOD_5(9). Dorr-Oliver have also done considerable development work on the oxygenator, the flow distribution system and the sand-biomass separation equipment.

The experimental work at the Water Research Centre is aimed at investigating the performance of a coupled anoxic-oxygenic fluidized bed system, operated in the pre-denitrification mode, for complete treatment (carbon and nitrogen removal) of municipal wastewater. The investigations have concentrated on the denitrification process(23); and the oxygenic reactor evaluation is in its early stages(10). The oxygenic reactor performance has been similar to that attained by others in terms of organic carbon removal. Low suspended solids removal efficiencies noted in the system appear to be related to poor oxygenator performance.

2.2 Nitrification in the Biological Fluidized Bed Process

The early investigations of nitrification by the fluidized bed process were related to ammonia removal from river water for potable supplies(3,5). Subsequently, Jeris et al(7,24) investigated the fluidized bed for separate stage nitrification of a secondary effluent from a conventional activated sludge process in a pilot-scale system identical to that used for carbon oxidation. As indicated by the performance data (Table 2.2), ammonia oxidation efficiencies approaching 99 percent were reported at a nominal hydraulic retention time of about 11 minutes. Bed volatile solids concentrations averaged 8,500 mg/L. The authors reported dissolved oxygen and alkalinity limitations which affected the process performance at peak NH_3-N loadings. The secondary effluent quality fed to the nitrification reactor was relatively poor, containing approximately 70 mg/L COD.

Dorr-Oliver Inc. investigated separate stage nitrification of secondary effluent as well as combined carbon oxidation-nitrification in a single fluidized bed. The separate stage investigation is being conducted at Greenwich and there are no published results available(27). The results on the pilot plant fluidized bed at Orillia show that combined carbonaceous BOD_5 removal and nitrification can be achieved in a single fluidized bed(25). At a nominal hydraulic retention time of 37 minutes, an organic loading of

TABLE 2.2 SUMMARY OF FLUIDIZED BED TREATMENT PERFORMANCE UNDER NITRIFICATION CONDITIONS*

Secondary effluent flow (L/min)	52
Recycle flow (L/min)	122
Hydraulic rentention time (min)	10.6
Temperature (°C)	24
Secondary effluent NH_3-N (mg/L)	19.1
Fluidized bed NH_3-N (mg/L)	0.2
% NH_3-N Removal	99
Secondary effluent organic-N (mg/L)	3.1
Fluidized bed effluent organic-N (mg/L)	2.3

*Jeris *et al*(7)

0.17 kg BOD_5/kg BVS•d and a temperature of 21°C, the wastewater NH_3-N concentration was reduced from approximately 15 mg/L to 0.4 mg/L. The median unfiltered effluent BOD_5 under these conditions was 18 mg/L.

3 EQUIPMENT AND PROCEDURES

3.1 Pilot Plant Description

The pilot-scale fluidized bed system used during the carbon oxidation phase is shown in Figure 3.1. The fluidized bed reactor consisted of a 28.9 cm diameter cylindrical column, 4.27 m tall with a conical entrance to improve inlet flow distribution. The expanded bed height was controlled at 3.9 m by a timer-actuated value. The total empty bed volume of the reactor to the wasting valve was 0.21 m^3.

The biomass from the wasted fraction of the bed was sheared from the support medium (quartzite sand) in an external system and the relatively clean medium returned to the column. During the early stages, a Wemco Model 6 Spiral Classifier was used for sand-biomass separation and the clean sand manually returned to the reactor. This system was replaced by a helical conveyor (Hapman Helix) which allowed mechanical sand return and appeared to improve separation. Optimization of the separation and return system was beyond the program. Excess biomass from the separator was collected in a storage tank and monitored routinely to define the biomass growth rate.

Pure oxygen was dissolved in the reactor feed through a proprietary oxygenator developed by Dorr-Oliver Inc. A proportional-integral digital feedback control loop on the oxygen supply allowed operation of the fluidized bed at any desired effluent dissolved oxygen concentration. Influent and effluent dissolved oxygen concentrations were monitored continuously with in-line Beckman dissolved oxygen probes. Effluent recycle ensured that the oxygen demand of the wastewater did not exceed the transfer capabilities of the oxygenator and allowed variation in the process hydraulic retention time (HRT) independent of the reactor hydraulic flux.

Temperature control of the reactor feed in the pilot plant was by means of immersion type heaters and coolers. A small sand-liquid separator operated at a high overflow rate was provided to recover any sand carried over from the reactor; this proved unnecessary, as there was never any significant loss of the support medium.

During the carbon oxidation phase the reactor feed was degritted domestic wastewater piped directly from the Burlington Skyway WPCP and settled in a 2.44 m diameter primary clarifier. During the nitrification studies, the reactor feed was nitrified secondary effluent from an extended aeration activated sludge plant at the Wastewater Technology Centre operated on degritted raw sewage from Burlington Skyway WPCP at a constant flow rate, but with diurnal variations in wastewater strength. The nitrified effluent from the activated sludge plant was fed directly to the fluidized bed. Ammonium chloride was added at a controlled rate to the feed to allow precise control of the system nitrogen loading. Sodium bicarbonate was also dosed at a controlled rate to prevent alkalinity limitations during the nitrification reactions. The quality of the fluidized bed feed during the various phases is discussed in detail later.

The support medium used throughout was quartzite sand with an effective size (d_{10}) of 0.48 mm and a uniformity coefficient (d_{60}/d_{10}) of 1.23. The particle size distribution is given in Figure 3.2.

3.2 Sampling and Analytical Procedures

3.2.1 Carbon Oxidation Mode. During the steady-state experimentation (Section 4), the performance of the system was assessed by daily compositing of 24 samples of reactor feed and of treated effluent taken over 24 hours. Samples were analyzed daily for filterable organic carbon (FOC), suspended solids (SS) and volatile suspended solids (VSS). Samples were analyzed three times per week for total and filterable chemical oxygen demand (COD), total and filterable Kjeldahl nitrogen (TKN), ammonia nitrogen (NH_3-N), and nitrate and nitrite nitrogen (NO_3-N and NO_2-N). Influent and effluent samples were analyzed weekly for total and filterable BOD_5. Once a week, treated effluent samples were settled for 30 minutes under quiescent conditions and the supernatant analyzed for total BOD_5, total COD and suspended solids.

The concentration of biomass in the reactor was measured three times a week by sampling at six locations in the bed and determining the mean volatile solids concentration. A special device designed to take a representative sample of the fluidized bed including sand, biomass and liquid was used. These samples were composited and volatilized to measure the mean bed volatile solids (BVS) concentration. Once a week, the bed samples were analyzed individually to measure the distribution of concentrations over the reactor profile. At the same time, samples of the reactor liquid were syphoned from the six BVS sampling locations and analyzed for TKN and FOC. Dissolved oxygen measurements were made in situ at each location.

During fluidized bed operation under diurnal flow and concentration conditions (Section 5), discrete hourly sampling of influent and effluent streams was conducted weekly in addition to the routine daily composite sampling. The 24-hourly samples were analyzed for total and filterable COD, FOC and suspended solids. The sampling and

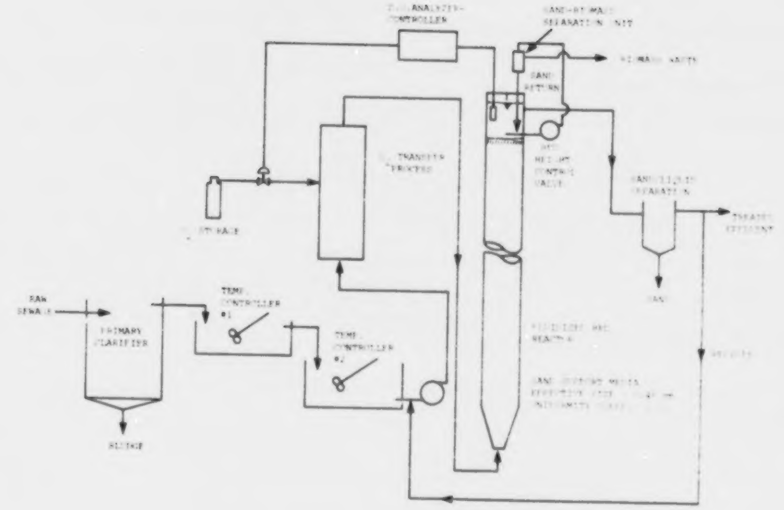

ANALYZER-
CONTROLLER
SEPARATION UNIT
SAND
RETURN
TRANSFER
STORAGE
HEIGHT
CONTROL
VALVE
SEPARATION
TREATED
EFFLUENT
SAND
RAW
SEWAGE
PRIMARY
CLARIFIER
TEMP.
CONTROLLER
#1
TEMP.
CONTROLLER
#2
SLUDGE
FLUIDIZED BED
REACTOR

FIGURE 3.1 BIOLOGICAL FLUIDIZED BED PROCESS FLOW

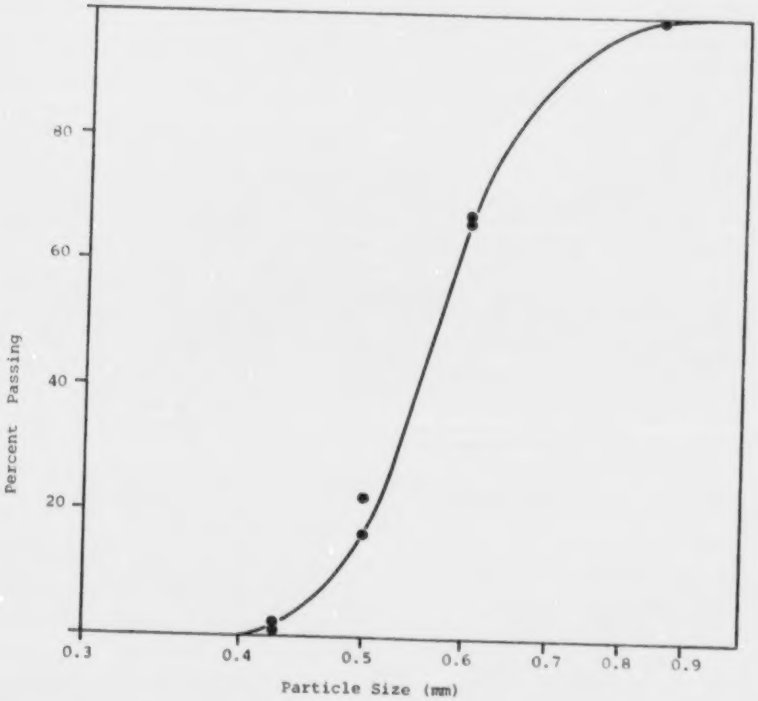


FIGURE 3.2 FLUIDIZED BED SAND PARTICLE SIZE DISTRIBUTION

analytical routine for assessment of dynamic performance (Section 6) was designed for the time series experimental conditions and is discussed in detail in another section.

All analyses were conducted by the Wastewater Technology Centre(28).

3.2.2 **Nitrification Mode.** Sampling procedures during the nitrification investigations were similar to those for the carbon oxidation studies; composite samples of influent and effluent from the fluidized bed, however, were analyzed daily for nitrogen compounds (TKN, NH_3-N, NO_2-N and NO_3-N) and alkalinity in addition to FOC and suspended solids. During the weekly reactor profile sampling, NO_3-N and NO_2-N were included in the analyses; FOC was omitted due to the low concentrations of degradable organic carbon in the feed.

3.3 Fluidized Bed Mixing Characteristics

Prior to the detailed evaluation of fluidized bed performance, tracer studies were done under different hydraulic loading and recycle conditions to define the mixing characteristics. An impulse input of Rhodamine WT was used as the tracer. The reactor operating conditions are summarized in Table 3.1. The tanks-in-series model described by Levenspiel(29) was used to assess the reactor mixing characteristics.

TABLE 3.1 EVALUATION OF FLUIDIZED BED MIXING CONDITIONS: EXPERIMENTAL CONDITIONS

	Tracer Study	
	1	2
Primary effluent flow (L/min)	21.24	7.08
Recycle flow (L/min)	42.48	21.25
Hydraulic flux (m/min)	0.97	0.43
Bed void volume (L)	158.6	166.4
Theoretic hydraulic retention time* (min)	7.47	23.5

*Based on primary effluent flow rate and the void volume of the reactor (total reactor volume minus volume occupied by sand).

Numerous authors have shown that plug flow hydraulics provide an adequate description of a fluidized bed reactor operated without recycle(2,8,30,31).

The relationship developed by Leva(32) to describe dispersion in liquid fluidized systems

$$D_L = 0.33 \, v \, D_p^{0.85} \tag{3.1}$$

where D_L is the diffusivity coefficient (ft^2/h),
v is the superficial fluid velocity (ft/h), and
D_p is the particle diameter (ft).

indicates that the reactor dispersion number ($D/\mu L$) for the pilot plant reactor at typical hydraulic fluxes (0.50 to 1.00 m/min) ranges from 0.003 to 0.006. This range of values is described by Levenspiel(29) as indicative of a small amount of dispersion, and readily approximated by a plug flow model.

When the reactor is operated under recycle conditions, the mixing characteristics can be treated as a continuous stirred tank reactor (CSTR)(33). This was confirmed by the tracer studies conducted at hydraulic fluxes of 0.97 m/min and 0.43 m/min and recycle ratios of 2:1 and 3:1, respectively. The external residence-time distributions for the two tracer studies are shown in Figures 3.3 and 3.4. In both cases, the tanks-in-series model fitted to the experimental data indicates that the mixing characteristics of the system approach a single ideal CSTR (Table 3.2). As the recycle ratio increases, the reactor mixing conditions more closely approximate an ideal CSTR. Under conditions typical of the carbon oxidation and nitrification studies, an ideal CSTR model adequately describes the system hydraulics.

TABLE 3.2 RESULTS OF FLOW MODELLING EXPERIMENTS

	Tracer Study	
	1	2
Hydraulic flux (m/min)	0.97	0.43
Recycle ratio (Q_R Q_I)	2.0	3.0
Theoretic hydraulic retention time (min)	7.47	23.5
Dye peak time (min)	3.00	4.08
Percent dye recovery	115.	105.
Number of CSTR's in series	1.67	1.21

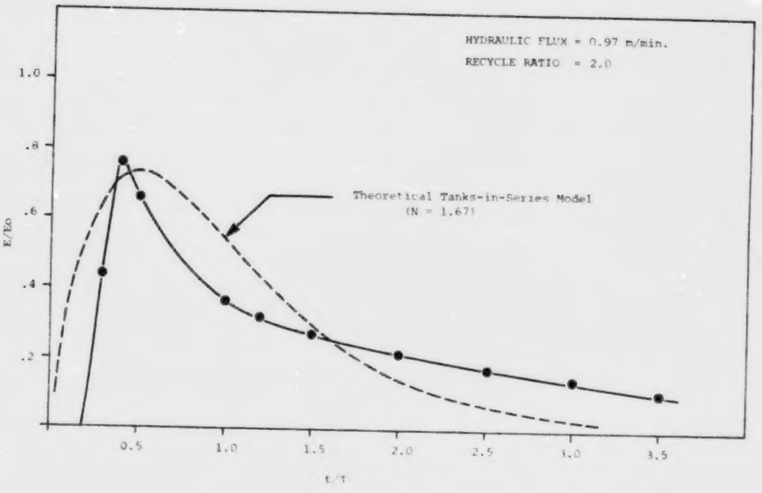


FIGURE 3.3 TRACER RESPONSE (RUN 1)

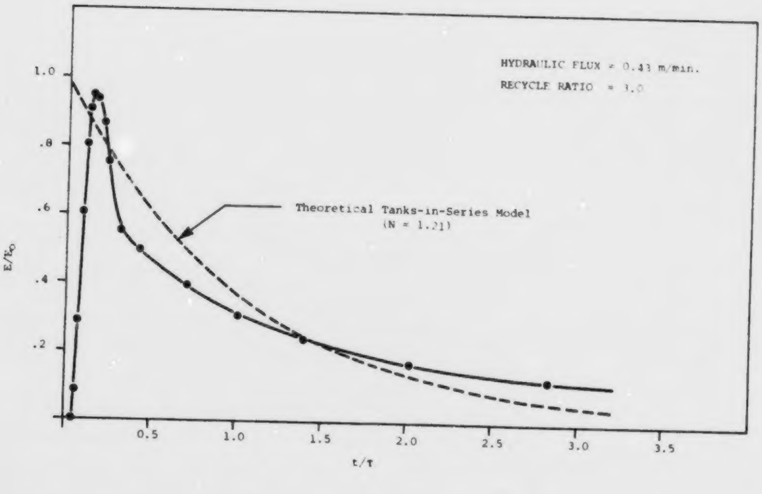


FIGURE 3.4 TRACER RESPONSE (RUN 2)

4 PSEUDO-STEADY STATE PERFORMANCE FOR CARBON OXIDATION

The objective of the pseudo-steady state investigations was to define the input variables affecting process performance and to assess the performance of the biological fluidized bed process over a range of hydraulic and substrate loading conditions. The input variables included influent substrate concentration (C_o), temperature (T), effluent dissolved oxygen concentration (DO), hydraulic retention time (HRT) and recycle ratio ($R = Q_R/Q_I$).

4.1 Experimental Program

4.1.1 Experimental Design. To assess the influence of five input variables (T, HRT, R, DO, C_o) on the fluidized bed process within the available time frame, a 2^{5-1} fractional factorial design(42,43) was developed to allow evaluation of each parameter at two levels within 16 experimental runs. An additional four replicate runs at the mid-range of each parameter were included to allow estimation of the pure experimental error. The ranges of the input variables were based on published information on the performance of the fluidized bed system in the carbon oxidation mode, primarily from Jeris *et al*(7). The levels at which the input variables were controlled are summarized in Table 4.1.

TABLE 4.1 CONTROL LEVELS FOR INPUT VARIABLES DURING PSEUDO-STEADY STATE EXPERIMENTS

VARIABLE	Level		
	-1	0	+1
Temperature (°C)	8	12	16
HRT (min)*	10	20	30
Recycle ratio (Q_R/Q_I)	2	2.5	3
Effluent DO (mg/L)	2	5	8
Substrate concentration (FOC, mg/L)	30-40	40-50	50-60

*Hydraulic Retention Time, based on primary effluent flow rate (Q_I) and empty bed reactor volume.

All input variables, except influent substrate concentration (C_o), could be readily controlled at the desired levels. To provide some degree of control on C_o, the fluidized bed was operated on stored sewage. Raw sewage from the Burlington Skyway WPCP was stored in mixed tanks at selected times of the day depending on the desired organic carbon (FOC) concentration, and each batch of stored sewage was used for 24 hours. Despite the storage procedure, it was difficult to consistently maintain the desired feed strength due to seasonal variations in the wastewater and inflow/infiltration problems in the sewerage system. The desired range of FOC concentration was 30 to 60 mg/L, but the actual concentration varied from 16 to 64 mg/L. The poor control over raw waste strength adversely affected the experimental design as discussed later.

In addition to the 20 experimental runs initially included in the design, five pseudo-steady state runs were conducted at a temperature of 20 to 22°C during the evaluation of effluent solids filterability (Section 7.2.3). The results were included in the data analysis.

4.1.2 **Pilot Plant Operation.** Each steady state condition was maintained for a period of seven days. The initial four days, after changes in the control variable set points, were used to acclimate the system. During the first three days, the system was fed directly from the raw sewage entering the Burlington Skyway WPCP and thus experienced typical diurnal variations in raw waste strength. Stored sewage feed was commenced on the fourth day and continued for the remainder of the run. Performance data were collected during the final three days.

The fluidized bed reactor performance was based on the average of three 24-hour composite influent and effluent samples collected during the final three days of operation.

In specific instances, the length of the experimental run or the length of the acclimation period were altered slightly. It was found that the fluidized bed could not be operated successfully for an extended period at a total hydraulic flux of 0.32 $m^3/m^2 \cdot min$ (equivalent to an HRT of 30 minutes and a recycle ratio of 2). At these conditions, the support media was poorly fluidized and it was difficult to control film thicknesses; experimental runs 11 and 12 and runs 13 and 14 were conducted without acclimation. Several experimental runs were duplicated due to mechanical or operational difficulties. Overall, 28 experimental runs were conducted.

The experimental conditions for each run are summarized in Table 4.2, showing the discrepancy between the designed influent substrate concentration (C_o, mg/L FOC) and the actual concentration attained by storage. The deviation of C_o from the

TABLE 4.2 PSEUDO-STEADY STATE EXPERIMENTAL CONDITIONS

	Designed Input Variables						
Run No.	C_o (mg/L)	Temp. (°C)	DO (mg/L)	HRT (min)	Recycle Ratio	Actual Avg. Influent FOC (mg/L)	Total Hydraulic Flux ($m^3/m^2 \cdot min$)
1	50 - 60	16	2	10	2	55	0.97
2	30 - 40	16	8	10	2	43	0.97
3	30 - 40	8	2	10	2	38	0.97
4	50 - 60	8	8	10	2	41	0.97
5	50 - 60	16	8	10	3	38	1.29
6	30 - 40	8	8	10	3	23	1.29
7	30 - 40	16	2	10	3	45	1.29
8	50 - 60	8	2	10	3	58	1.29
9	40 - 50	12	5	20	2.5	49	0.57
10	40 - 50	12	5	20	2.5	50	0.57
11-1	50 - 60	8	2	30	2	52	0.32
11-2	50 - 60	8	2	30	2	48	0.32
12	30 - 40	8	8	30	2	30	0.32
13	30 - 40	16	2	30	2	19	0.32
14-1	50 - 60	16	8	30	2	62	0.32
14-2	50 - 60	16	8	30	2	30	0.32
15	30 - 40	8	2	30	3	22	0.43
16	30 - 40	16	8	30	3	27	0.43
17-1	50 - 60	16	2	30	3	78	0.43
17-2	50 - 60	16	2	30	3	38	0.43
18	50 - 60	8	8	30	3	46	0.43
19	40 - 50	12	5	20	2.5	34	0.57
20	40 - 50	12	5	20	2.5	27	0.57
21	40 - 50	20	5	20	2.5	35	0.57
22	40 - 50	20	5	20	2.5	37	0.57
23	40 - 60	20	5	10	2	30	0.97
24	30 - 40	20	5	20	2.5	27	0.57
25	30 - 40	20	5	30	3	28	0.43

experimental design level, and the necessity to eliminate the results of some experimental runs due to mechanical and operational problems, created difficulties in the data analysis as the resultant experimental design was not orthogonal.

4.1.3 Wastewater Characteristics. As the pilot plant was operated on stored raw sewage collected at times which would provide some control of the influent substrate concentration, the specific concentration of the feed during a particular run is more relevant in the evaluation of performance than is the mean concentration over the entire period. However, the overall data indicate the range of feed quality under which the process was evaluated. The characteristics of the feed are summarized in Table 4.3, based on the data presented in Figure 4.1 (suspended solids and filterable organic carbon) and Figure 4.2 (total and filterable BOD_5).

TABLE 4.3 REACTOR FEED QUALITY STEADY STATE CONDITIONS

Parameter	Concentration (mg/L) Mean	Standard Deviation	90%*	Number of Observations
Total BOD_5	104	39.1	150	27
Filterable BOD_5	51	24.2	88	27
Filterable organic carbon	39	13.9	57	78
Suspended solids	113	30.4	144	77

*90% refers to percent of observations equal to or less than stated value.

4.2 Experimental Results

4.2.1 Effect of Input Variables on Process Performance. The correlation coefficient between the process response and the controlled input variables (main effects and two factor interactions) were computed to determine the factors significantly influencing the performance of the biological fluidized bed.

Figure 4.3 shows the correlations between removal rates in terms of total and filterable BOD_5 and suspended solids (g/g BVS•d). In all three cases, the specific removal rate was significantly correlated to the influent substrate concentration (C_o), the process hydraulic retention time (HRT) and the concentration-hydraulic retention time interaction (C_o•HRT). This two-factor interaction is related to the process substrate loading. There was also a statistically significant (95 percent confidence) correlation between the total BOD_5 removal rate and the interaction between concentration and temperature (C_o•T) and between the suspended solids removal rate and the interaction between the temperature and hydraulic retention time (T•HRT). However, these two-factor interactions appear to have no physical significance.

The relationship between treated effluent quality, in terms of the concentration of total and filterable BOD_5 and suspended solids, and the controlled input

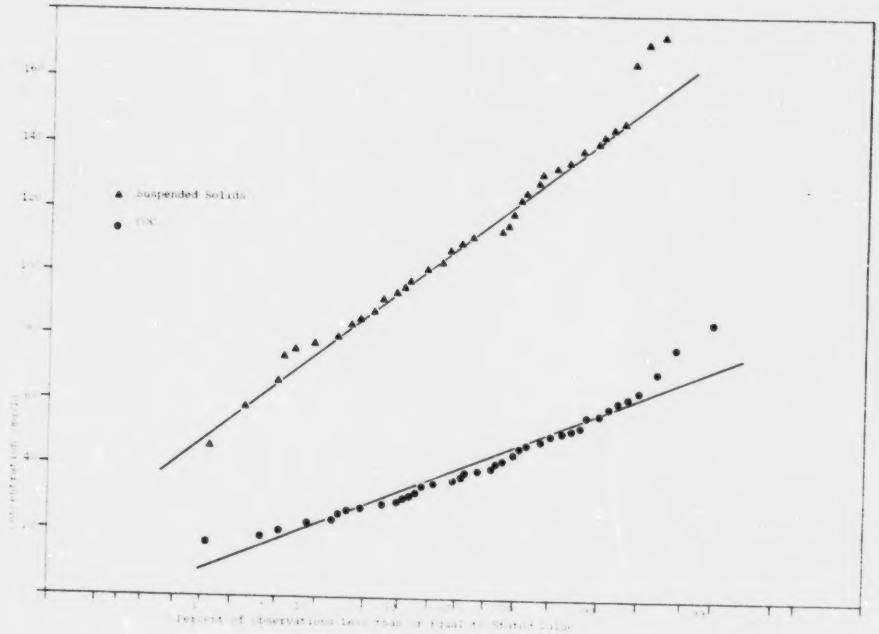


FIGURE 4.1 REACTOR FEED QUALITY - FILTERABLE ORGANIC CARBON AND SUSPENDED SOLIDS

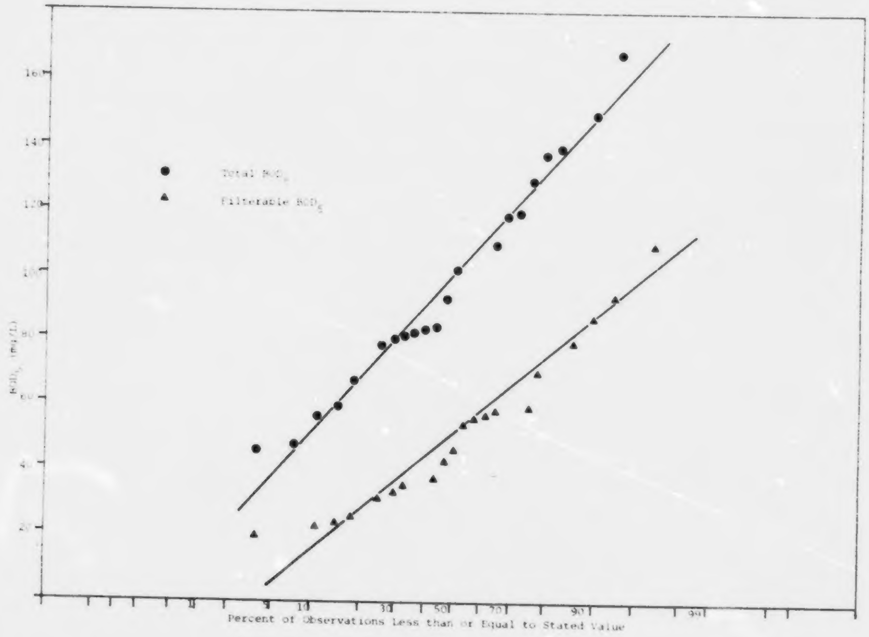
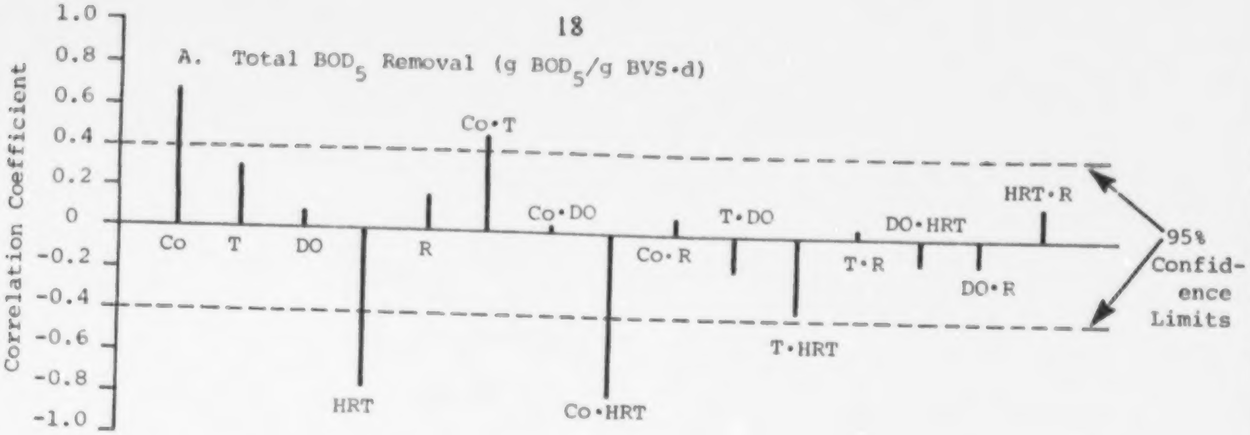
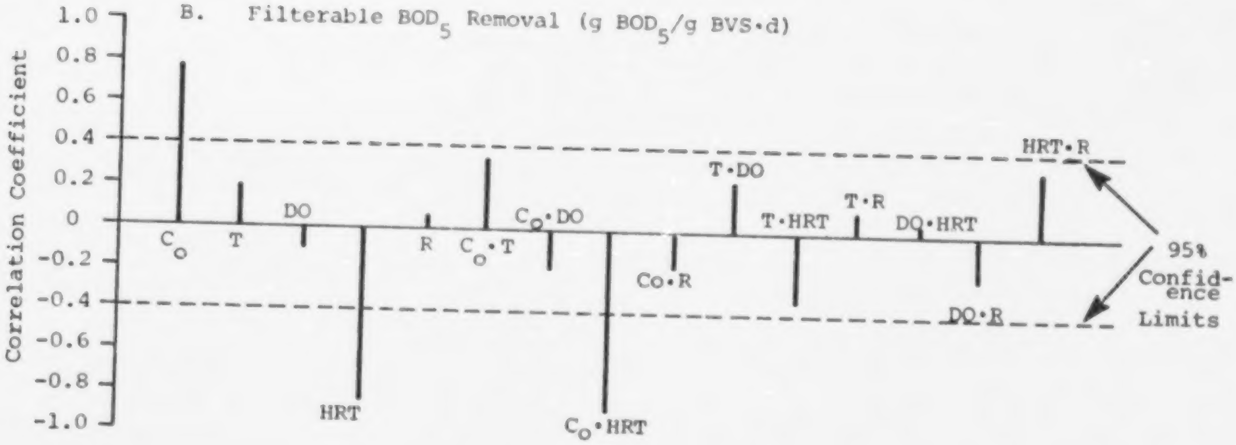


FIGURE 4.2 REACTOR FEED QUALITY - TOTAL AND FILTERABLE BOD5


A. Total BOD_5 Removal (g BOD_5/g BVS•d)


B. Filterable BOD_5 Removal (g BOD_5/g BVS•d)

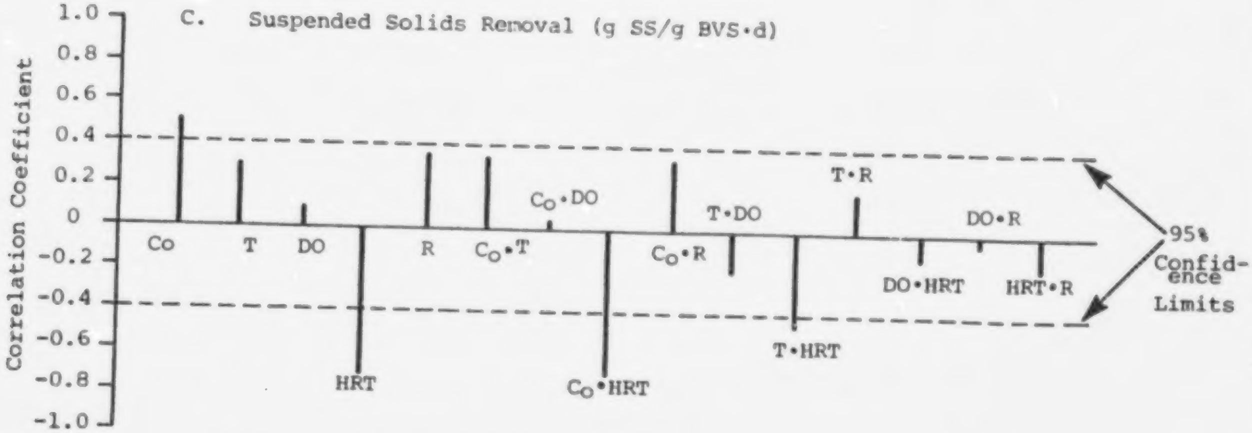

C. Suspended Solids Removal (g SS/g BVS•d)

FIGURE 4.3 CORRELATION BETWEEN SPECIFIC REMOVAL RATES AND INPUT VARIABLES

variables are shown in Figure 4.4. As in Figure 4.3, the influent concentration, hydraulic retention time and process loading (C_o•HRT) significantly influence the effluent quality. In addition, the effluent total and filterable BOD_5 concentration appears to be correlated to the effluent dissolved oxygen concentration. Jeris et al(7) concluded, based on column dissolved oxygen concentration profiles, that there was an oxygen limitation when the reactor dissolved oxygen concentration was less than 2 to 4 mg/L. The data also indicate a negative correlation between effluent dissolved oxygen concentration and effluent quality, indicative of improved process performance at a higher effluent dissolved oxygen concentration. However, the impact of effluent dissolved oxygen concentration is relatively minor compared to the effect of substrate concentration, process loading and system HRT.

Based on these data, there is little advantage in operating at high effluent dissolved oxygen concentrations and there are significant cost disadvantages related to maintaining an excessively high effluent dissolved oxygen concentration.

In addition to the relationship between effluent suspended solids concentration and influent suspended solids concentration, HRT and suspended solids loading, there was also a significant effect due to the interaction of HRT and recycle ratio. This two-factor interaction is related to the process hydraulic flux. The impact was relatively minor. Neither temperature nor recycle ratio had a significant effect on performance within the ranges investigated.

Based on the pseudo-steady state data, linear empirical models were developed relating the specific removal rates (g/g•d) and the process effluent quality to the major variables influencing performance. (See Table 4.4). It is evident that the parameter estimates are poorly defined and only the parameter associated with the loading term (β_3) is consistently different from zero at the 95 percent confidence level. Poor definition of the parameter estimates is a direct result of the problems associated with maintaining the experimentally desired influent substrate concentration.

4.2.2 Effect of Mass Loading on Process Removal Rate. The effect of process mass loading, in terms of total BOD_5, filterable BOD_5 and suspended solids, on the specific mass removal rate (g/g BVS•d) are shown in Figures 4.5, 4.6 and 4.7 respectively.

The total BOD_5 and filterable BOD_5 removal efficiencies were relatively constant at about 60 to 70 percent. From these data, there was no indication that the maximum removal rate for organic carbon had been attained. Figures 4.5 and 4.6 show

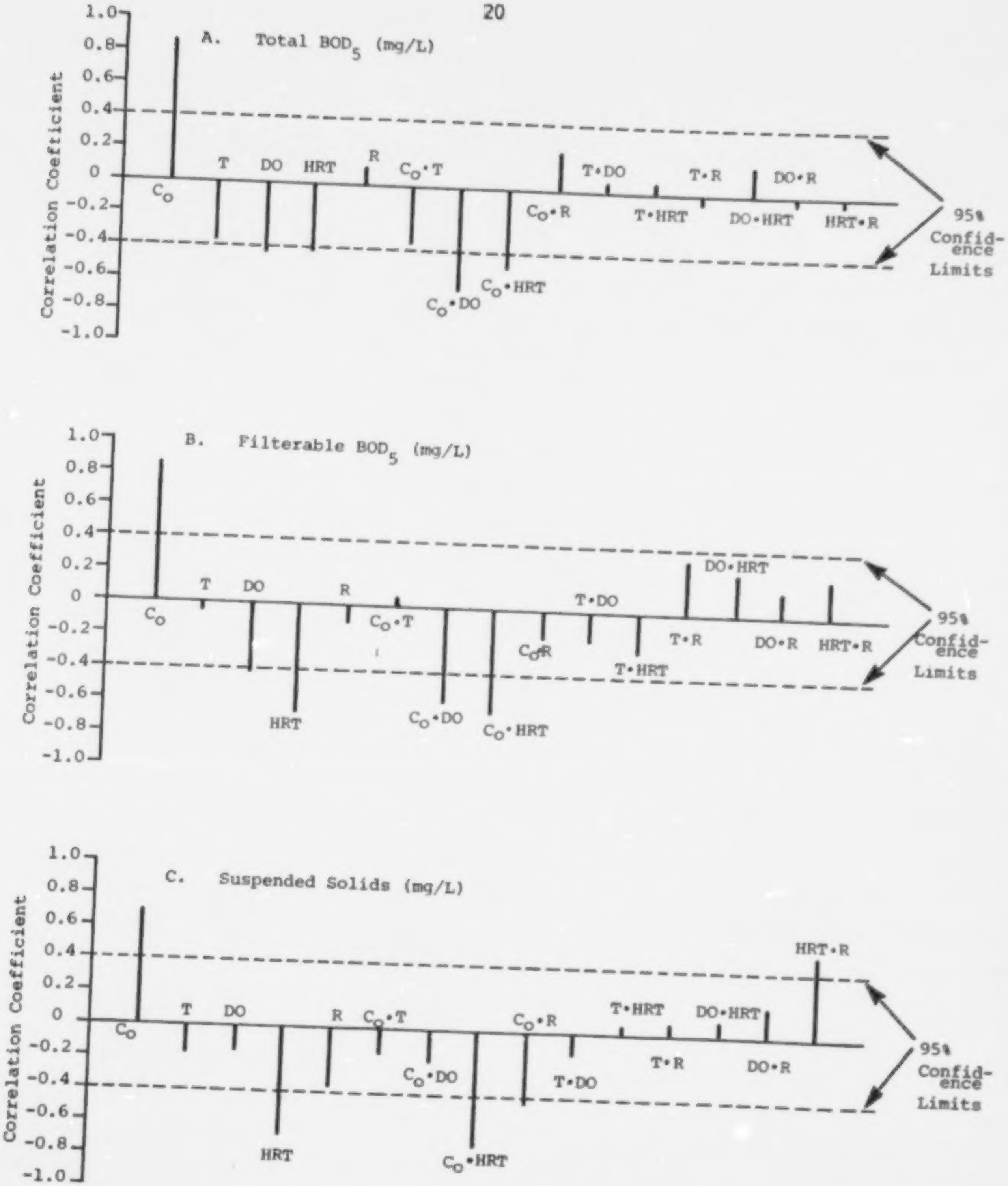


FIGURE 4.4 CORRELATION BETWEEN TREATED EFFLUENT CONCENTRATION AND INPUT VARIABLES

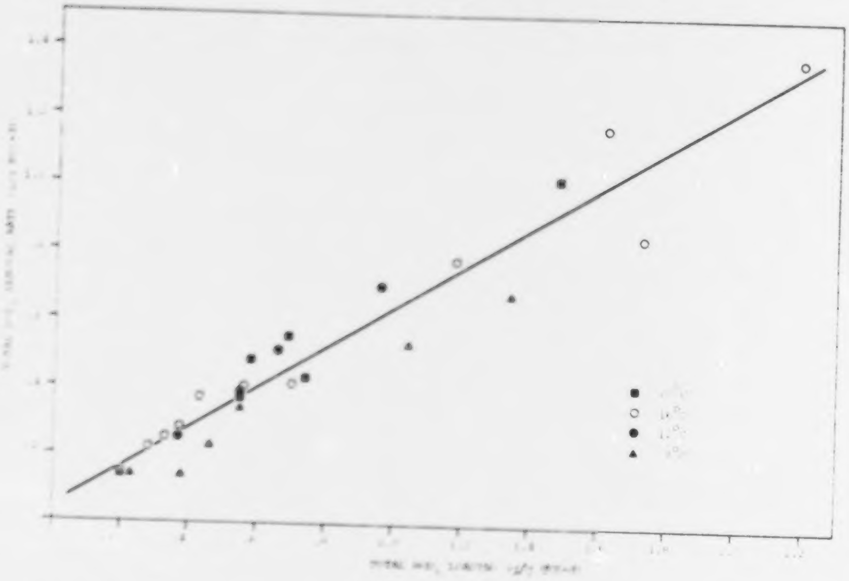


FIGURE 4.5 SPECIFIC REMOVAL RATE - TOTAL BOD_5

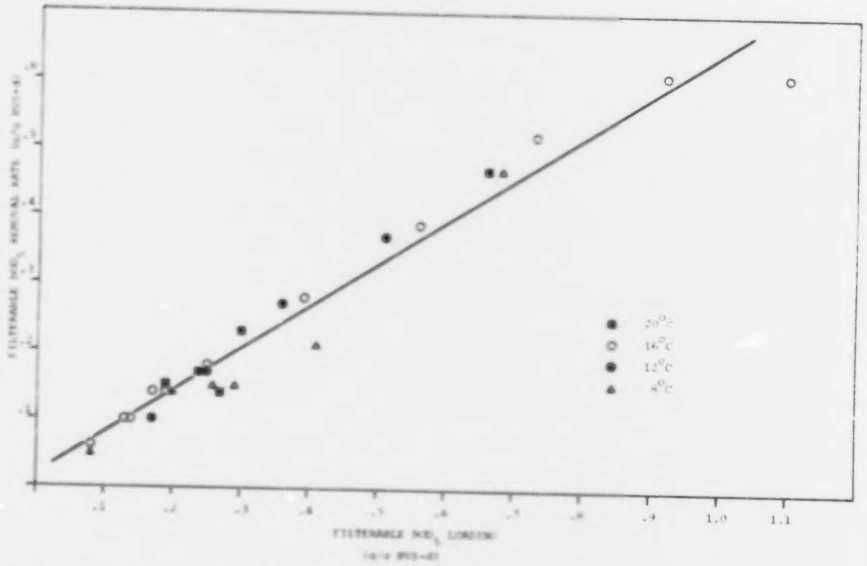


FIGURE 4.6 SPECIFIC REMOVAL RATE - FILTERABLE BOD_5

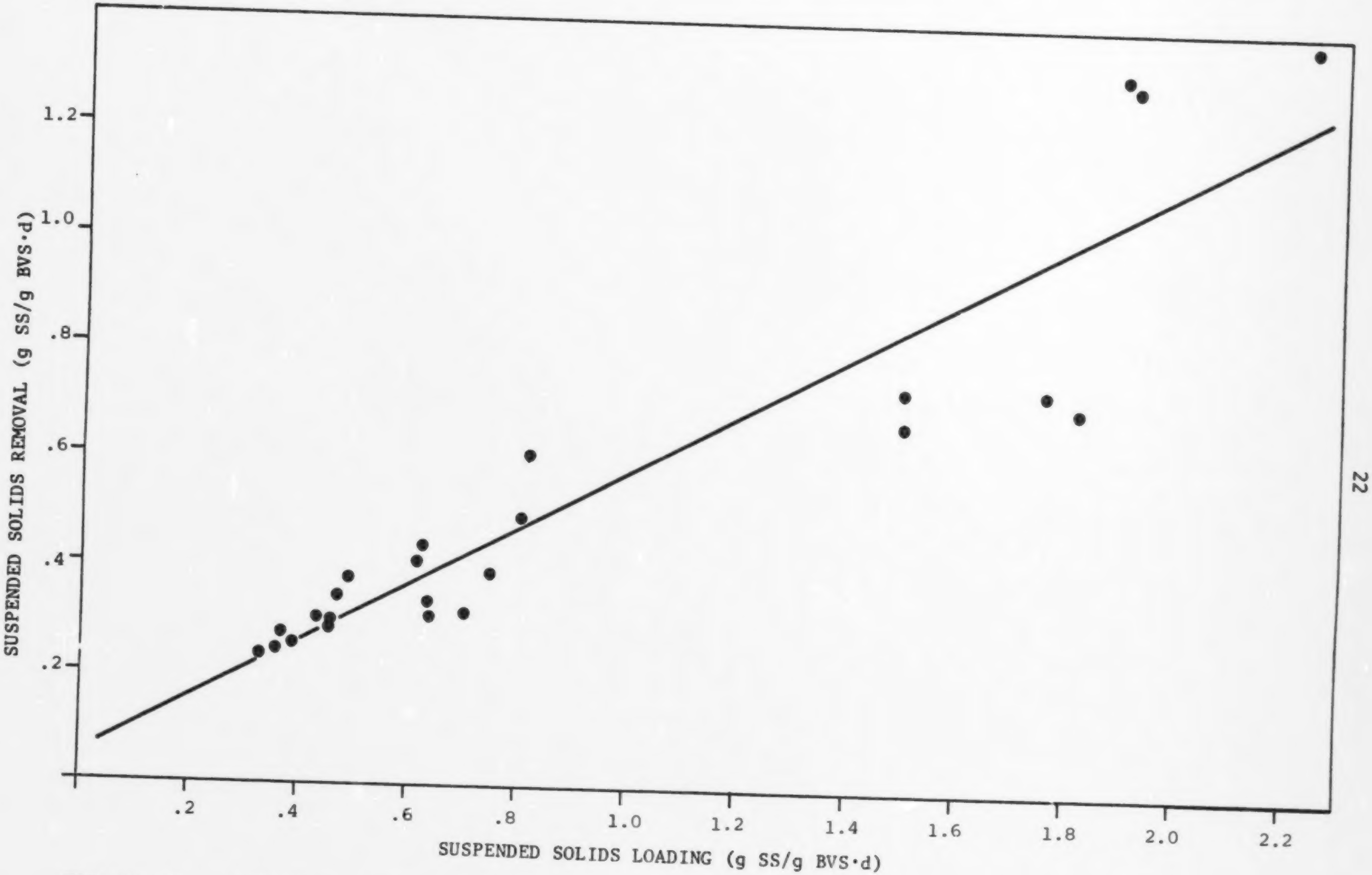


FIGURE 4.7 SPECIFIC REMOVAL RATE - SUSPENDED SOLIDS

TABLE 4.4 PSEUDO-STEADY STATE REGRESSION MODELS

Output Response, Y	Parameter Estimates	95% Confidence Interval	Model
Total BOD_5 Removal Rate (g/g•d)	$\beta_o = 0.11$	$\beta_o \pm 0.33$	Y + 0.30 + 0.0013 C_o -0.0095 HRT + 0.031 L
	$\beta_1 = 0.0013$	$\beta_1 \pm 0.0056$	
	$\beta_2 = -0.095$	$\beta_2 \pm 0.27$	
	$\beta_3 = 0.031$	$\beta_3 \pm 0.00005$	
Filt. BOD_5 Removal Rate (g/g•d)	$\beta_o = 0.034$	$\beta_o \pm 0.061$	Y = 0.080 + 0.00059 C_o -0.0023 HRT + 0.043 L
	$\beta_1 = 0.00059$	$\beta_1 \pm 0.0022$	
	$\beta_2 = -0.023$	$\beta_2 \pm 0.054$	
	$\beta_3 = 0.043$	$\beta_3 \pm 0.021$	
Eff. Total BOD_5 Concentration (mg/L)	$\beta_o = 2.54$	$\beta_o \pm 19.2$	Y = 0.88 + 0.10 C_o + 0.17 HRT + 2.28 L
	$\beta_1 = 0.10$	$\beta_1 \pm 0.33$	
	$\beta_2 = 1.71$	$\beta_2 \pm 15.6$	
	$\beta_3 = 2.28$	$\beta_3 \pm 3.01$	
Eff. Filt. BOD_5 Concentration (mg/L)	$\beta_o = 0.83$	$\beta_o \pm 5.48$	Y = 1.51 + 0.20 C_o -0.034 HRT + 1.07 L
	$\beta_1 = 0.20$	$\beta_1 \pm 0.20$	
	$\beta_2 = -0.34$	$\beta_2 \pm 4.80$	
	$\beta_3 = 1.07$	$\beta_3 \pm 1.86$	
Eff. Susp. Solids Concentration (mg/L)	$\beta_o = -6.09$	$\beta_o \pm 19.14$	Y = -13.01 + 0.24 C_o + 0.346 HRT + 2.77 L
	$\beta_1 = 0.24$	$\beta_1 \pm 0.29$	
	$\beta_2 = 3.46$	$\beta_2 \pm 17.71$	
	$\beta_3 = 2.77$	$\beta_3 \pm 2.91$	

Model Form:

$$Y = \beta_o + \beta_1 C_o + \beta_2 \left(\frac{HRT - 20}{10}\right) + \beta_3 L$$

where: C_o is influent substrate concentration (mg/L)
HRT is hydraulic retention time (minutes)
L is process substrate loading ($kg/m^3 \cdot d$)

the lack of temperature dependency in the carbon oxidation reaction. This is not unexpected as the maximum removal rate was not attained.

The linear relationship between mass BOD_5 loading and mass removal indicates that, in the fluidized bed process, the percent removal of BOD_5 is independent of process loading. This observation is in agreement with the results presented by Murphy and Wilson(34) for the rotating biological contractor (RBC) process.

Similarly, the data (Figure 4.7) for suspended solids mass loading and removal indicate that removal efficiency was relatively constant at between 55 and 60 percent up to a suspended solids loading of about 1.0 g SS/g BVS•d. At higher loadings, there was considerable variability.

4.2.3 Effect of Mass Loading on Effluent Quality. The effect of the process' total BOD_5 loading (g/g BVS•d) on the effluent quality in terms of total and filterable BOD_5 is shown in Figure 4.8. The influence of suspended solids removal efficiency is evident. The process could produce a treated effluent containing less than 15 mg/L filterable BOD_5 at a process loading of up to approximately 1.0 g total BOD_5/g BVS•d. However, at this level, the effluent suspended solids from the process added 20 mg/L BOD_5 to the final effluent. The concentration of suspended solids at this loading, based on Figure 4.9, was approximately 50 mg/L. To attain an effluent total BOD_5 concentration of 15 mg/L or less from the fluidized bed process necessitates operation at a total BOD_5 mass loading of less than 0.1 g/g BVS•d. Effluent quality is limited by suspended solids removal rather than the ability to effect carbon oxidation.

The performance of the fluidized bed pilot plant is compared with the results from pilot plants at Orillia(25) and Nassau County(7) in Table 4.5. The organic loadings (g BOD_5/g BVS•d) necessary for an effluent quality of 20 mg/L BOD_5 were comparable at Burlington and Orillia. Jeris et al(7) reportedly achieved this level of treatment efficiency at a considerably higher organic loading. The improvement in BOD_5 removal efficiency at Nassau County was related to the relatively low influent suspended solids concentration and a proportionally lower contribution of effluent suspended solids to the effluent total BOD_5.

Both Orillia and Nassau County results indicated slightly better suspended solids removal than attained at Burlington (Figure 4.10). The removal of suspended solids may be related to the biofilm thickness(27). The pilot plant at Orillia was operated with relatively small sand inventories, resulting in the growth of thicker biofilms which may have improved suspended solids capture.

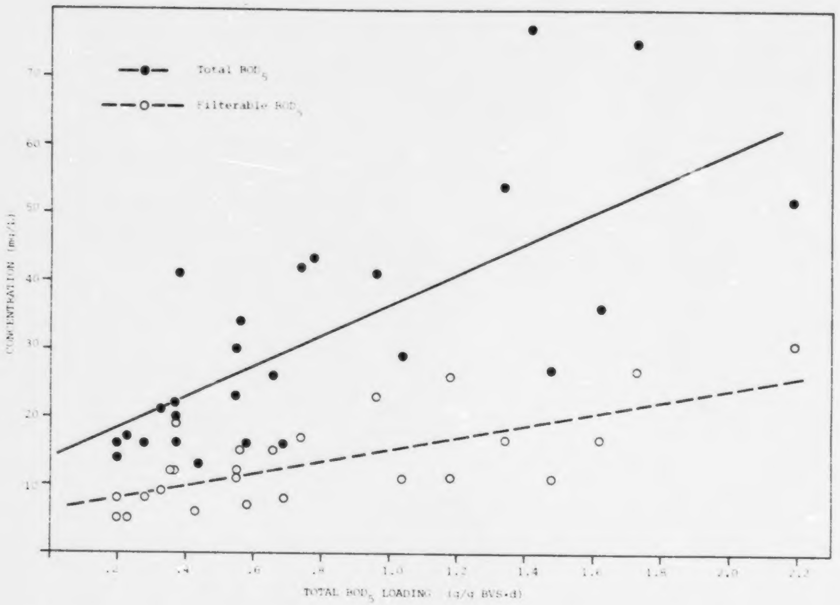


FIGURE 4.8 EFFECT OF PROCESS BOD$_5$ LOADING ON EFFLUENT BOD$_5$ CONCENTRATION

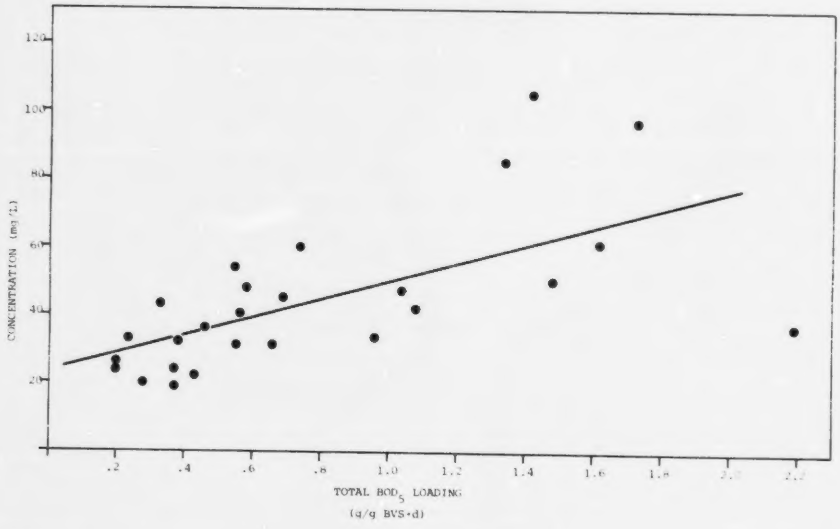


FIGURE 4.9 EFFECT OF PROCESS BOD$_5$ LOADING ON EFFLUENT SUSPENDED SOLIDS CONCENTRATION

TABLE 4.5 COMPARISON OF LOADING CONDITIONS NECESSARY TO PRODUCE A TREATED EFFLUENT CONTAINING 20 mg/L BOD_5 AND 30 mg/L SUSPENDED SOLIDS

Reference	Reactor Volume (m^3)	Wastewater Characteristics (mg/L)		Loading Conditions		
		BOD_5	SS	Organic (g BOD_5/g BVS•d)	Solids (g SS/g BVS•d)	Volume (kg BOD_5/m^3•d)
This study	0.21	104	113	0.27	0.28	3
Sutton et al(25)	10.4	67	104	0.25	0.40	3
Jeris et al(7)	0.95	94	65	0.62	0.43	8

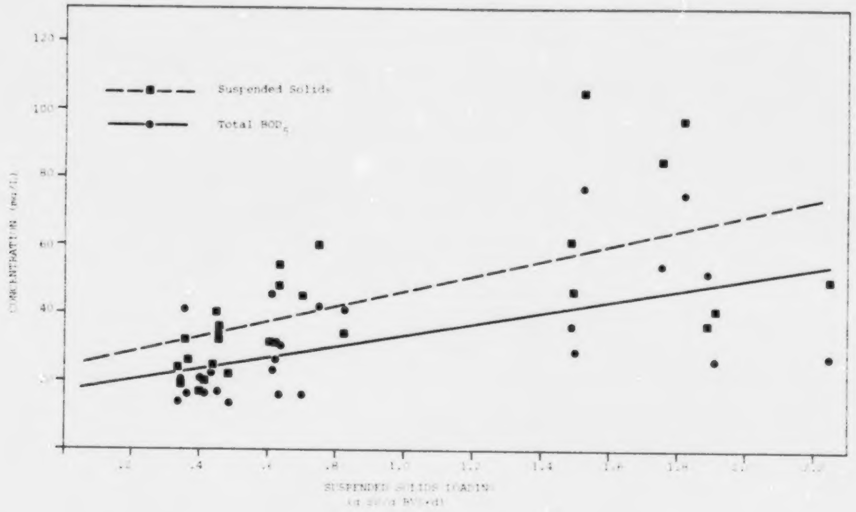


FIGURE 4.10 EFFECT OF PROCESS SOLIDS LOADING ON EFFLUENT QUALITY

Based on the pseudo-steady state operating data, the design fluidized bed loadings to attain a range of effluent quality objectives are summarized in Table 4.6. It is evident that this process can attain an effluent BOD_5 objective of 30 mg/L with considerably reduced reactor size relative to conventional systems, due to the high removal efficiency per unit volume. Much of the size advantage is lost when the effluent BOD_5 objective is reduced to 15 mg/L, due to the contribution of effluent suspended solids to the effluent total BOD_5. An effluent containing less than 15 mg/L suspended solids was not consistently achieved at the Burlington pilot plant under the range of operating conditions evaluated. Improving treatment efficiency to meet more stringent effluent quality requirements (15 mg/L BOD_5 and 15 mg/L suspended solids) is limited by the ability of process to effect suspended solids removal.

Because of the importance of residual suspended solids in the treated effluent, considerable additional effort was spent characterizing these suspended solids. The results are presented in Section 7.

4.2.4 Bed Volatile Solids Characteristics. The high concentration of biomass reported in other fluidized bed studies was confirmed at Burlington. The distribution of bed volatile solids concentrations resulting from the pseudo-steady state runs is shown in Figure 4.11. The mean concentration of volatile solids in the biological reactor was

TABLE 4.6 DESIGN FLUIDIZED BED LOADINGS TO ACHIEVE EFFLUENT QUALITY OBJECTIVES

Objective	Organic Loading	
	g BOD_5/g BVS•d	kg BOD_5/m^3•d
Total $BOD_5 \leq 30$ mg/L	0.71	7.7
Total $BOD_5 \leq 15$ mg/L	0.04	0.4
Filterable $BOD_5 \leq 15$ mg/L	0.97	10.5

Note: 1 kg/m^3•d = 62.5 lb/1000 ft^3•d

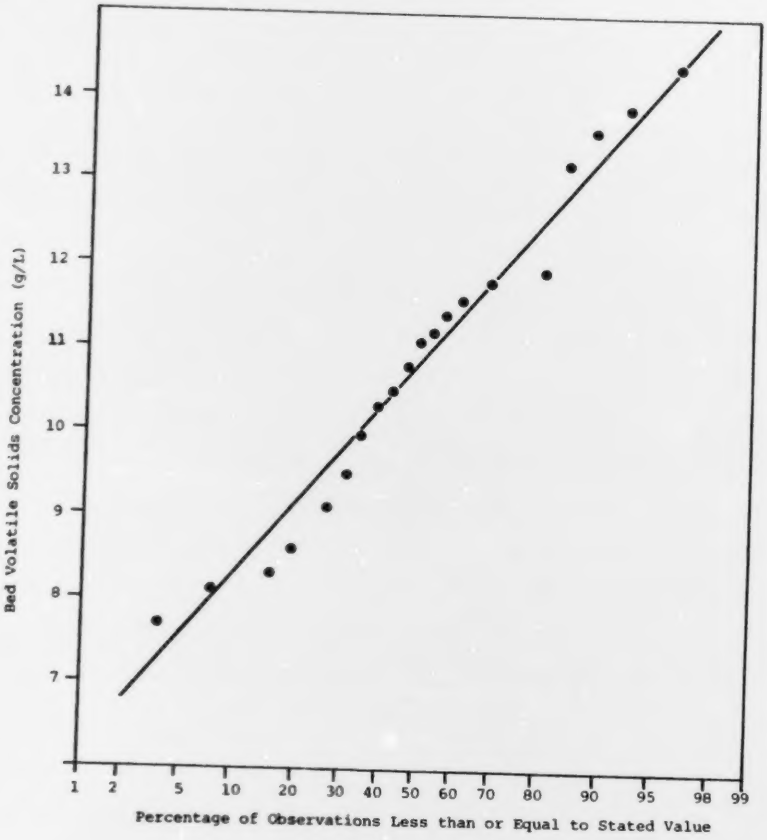


FIGURE 4.11 DISTRIBUTION OF BED VOLATILE SOLIDS CONCENTRATIONS

approximately 10.8 g/L. The high concentration of biomass which could be maintained in the reactor is directly responsible for the high loadings at which the fluidized bed can be operated. Although the biomass loading rates for the system are comparable to those for a conventional activated sludge process treating municipal wastewaters (0.25 g BOD_5/g VSS•d), the order-of-magnitude increase in the bed volatile solids concentration translates to a similar increase in the process loading relative to a conventional system designed to operate at approximately 0.45 kg $BOD_5/m^3 \cdot d^{(41)}$.

Reactor profile sampling indicated a distribution of bed volatile solids concentrations with depth in the reactor column. The results are shown as a probability plot in Figure 4.12. The region of the fluidized bed above a depth of approximately 1.2 m had a relatively uniform concentration of biomass, although the concentration at the top of the bed (3.7 m) was slightly lower than that in the middle (2.4 and 3.0 m). The volatile solids concentration in the most turbulent bottom region (0.6 m) was significantly lower than in the upper regions. A noticeable interface was evident in the bed. The biofilms below this interface were very thin ($\leq$50 μ) compared to thicknesses above (100 to 200 μ).

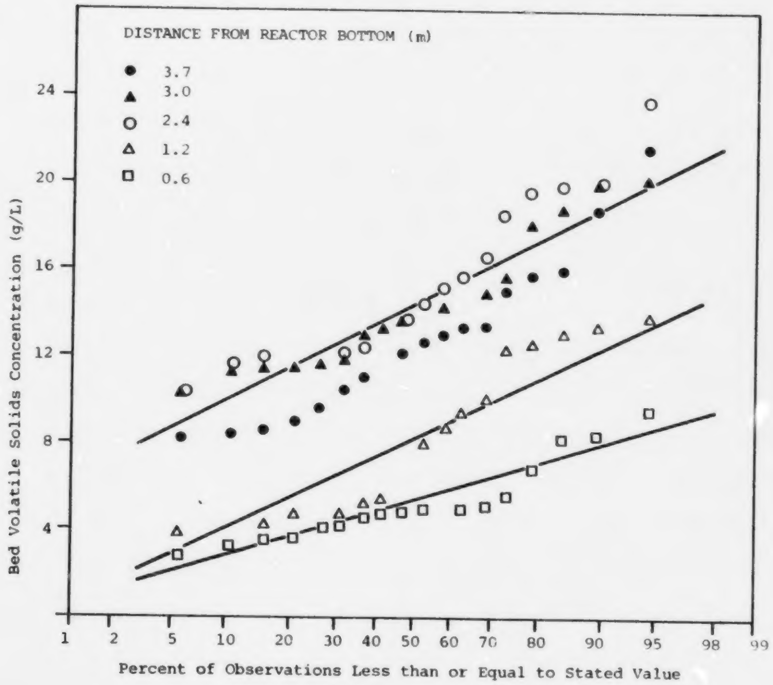


FIGURE 4.12 DISTRIBUTION OF VOLATILE SOLIDS CONCENTRATION WITH REACTOR HEIGHT

The position of the interface in the reactor varied depending on the hydraulic flux. The wide variation in the volatile solids concentration at 1.2 m from the bottom indicates that under some conditions the interface was above this. The interface was highest at the lowest flux conditions (0.32 $m^3/m^2 \cdot min$), where the support media was poorly fluidized and the high level of shear in the bottom of the reactor produced very thin films up to about 1.8 m from the inlet. Very thick films (>500 µ) were evident above the interface.

Profile sampling in the Dorr-Oliver pilot plant in Orillia shows a similar distribution of bed solids concentrations(25). However, the bed volatile solids were relatively constant above 0.99 m from the reactor entrance; and even at 0.38 m, a median BVS concentration of approximately 12.8 g/L was achieved at a hydraulic flux of 0.34 $m^3/m^2 \cdot min$. Dorr-Oliver expended considerable effort to improve the inlet flow distribution to the reactor to achieve uniform biomass concentrations. In addition, the sand used at Orillia was slightly smaller (d_{50} = 0.50 mm compared to d_{50} = 0.57 mm at Burlington). The smaller particle size also improves the fluidization properties under low flux conditions.

The improvement in suspended solids removal noted at Orillia (Section 4.2.3) may be related to the increased biofilm thicknesses in the bottom of the reactor resulting from the improved flow distribution. No noticeable improvement in carbon oxidation efficiency resulted from the improved seeding of the support medium.

4.2.5 Solids Retention Time and Solids Yield. The solids retention time (SRT) and solids yield (Y) for the fluidized bed process depend on the nature of the effluent suspended solids from the reactor. In a conventional system, effluent solids are considered to be biomass from the biological reactor and included in the calculated SRT. However, in the fluidized bed process, a significant quantity (approximately 65 percent depending on operating conditions) of the effluent solids represent feed solids not removed in the fluidized bed. (See Section 7.1.) A similar conclusion has been drawn by Sutton et al(25). Therefore, in the calculation of the process SRT for the fluidized bed, only the fraction of biomass solids contained in the final effluent should be included as biological solids lost from the process according to:

$$SRT_a = \frac{\text{volatile solids in the bioreactor}}{\text{(volatile solids intentionally wasted)} + 0.35 \text{ (volatile solids in effluent)}} \qquad (4.1)$$

This was the approach used in calculating the actual process SRT (SRT_a) from the pseudo-steady state results. The solids yield (Y_a g VS/g BOD_5 *removed) for the fluidized bed* was calculated from the total BOD_5 removed across the system. This is explained in more detail in Section 5.4.

The calculated SRT and yield coefficients for the pseudo-steady state runs are summarized in Table 4.7. The data from experimental runs 11 through 14 are not considered to represent steady state conditions. These were conducted at a relatively low flux (0.32 $m^3/m^2 \cdot min$) and operation at these flux conditions resulted in unstable biofilm conditions in the system (Section 4.1.2). The very high and variable solids yields calculated for these runs indicate that the process had not attained steady state.

At high organic loadings (F/M >1.0 g/g•d), the solids yield was high, with the average yield approximately 0.92 g VS/g BOD_5. At loadings more representative of typical operating conditions (0.2 g/g•d $\leq$ F/M $\leq$ 0.8 g/g•d), the solids yield averaged 0.69 g VSS/g BOD_5, comparable to those for conventional systems. However, longer-term data at constant operating conditions provide a better indication of the process solid yield. The data in Section 5.4 are more indicative of typical process yield.

There was no specific attempt to achieve a high concentration of waste biomass from the sand-biomass separation system. The wastage system was timer-actuated. The frequency and duration of the wasting operation was not specifically set to maximize waste sludge concentration. Dorr-Oliver(25), the Water Research Centre in Stevenage(44) and Ecolotrol(45) have all successfully improved the sand-biomass equipment to increase the waste solids concentration.

Although no optimization of waste sludge concentration was included at Burlington, the waste sludge solids content was comparable to that achieved from the secondary clarifier in a conventional suspended growth biological reactor. As shown in Figure 4.13, the median waste sludge concentration was approximately 1.15 percent VSS and concentrations as high as 2.4 percent VSS were achievable. Gravity thickening of the waste sludge increased the solids content by 100 to 200 percent.

4.2.6 **Oxygen Utilization.** The oxygen utilized to oxidize the wastewater BOD_5 was estimated from continuous records of influent and effluent dissolved oxygen concentrations during a number of the pseudo-steady state runs. Data from measurements on eight occasions are summarized in Table 4.8. The oxygen utilization rate will vary depending on the reactor loading and the rate of degradation of the particulate BOD_5 captured by the

TABLE 4.7 SOLIDS RETENTION TIME AND SOLIDS YIELD IN BIOLOGICAL FLUIDIZED BED SYSTEM

Run No.	Mean BVS (g/L)	Organic Loading (F/M, g BOD_5/g BVS•d)	SRT_a (days)	Y_a (g VS/g BOD_5)
1	11.91	1.73	1.4	0.82
2	11.42	1.62	1.6	0.53
3	11.14	1.42	1.7	1.38
4	11.78	1.34	1.6	0.93
5	9.53	1.18	1.4	0.89
6	8.13	1.04	1.8	1.08
7	9.16	2.19	1.0	0.74
8	8.57	3.52	1.1	1.32
9	11.21	0.96	2.1	0.67
10	11.88	0.66	4.1	0.48
11-1	14.36	0.46	4.2	1.06
11-2	13.89	0.38	2.2	1.89
12	13.60	0.20	2.3	3.11
13	11.13	0.20	5.7	1.26
14	11.90	0.33	5.8	0.68
15	9.48	0.23	6.2	1.14
16	9.22	0.43	5.2	0.53
17-1	11.58	0.70	2.7	0.89
17-2	10.27	0.56	3.7	0.67
18	10.53	0.55	4.6	0.64
19	13.17	0.37	4.5	0.89
20	10.80	0.55	4.7	0.54
21	9.96	0.74	3.0	0.78
22	11.75	0.55	3.4	0.81
23	8.32	1.48	1.6	0.61
24	8.29	0.69	3.7	0.49
25	7.70	0.58	4.5	0.46

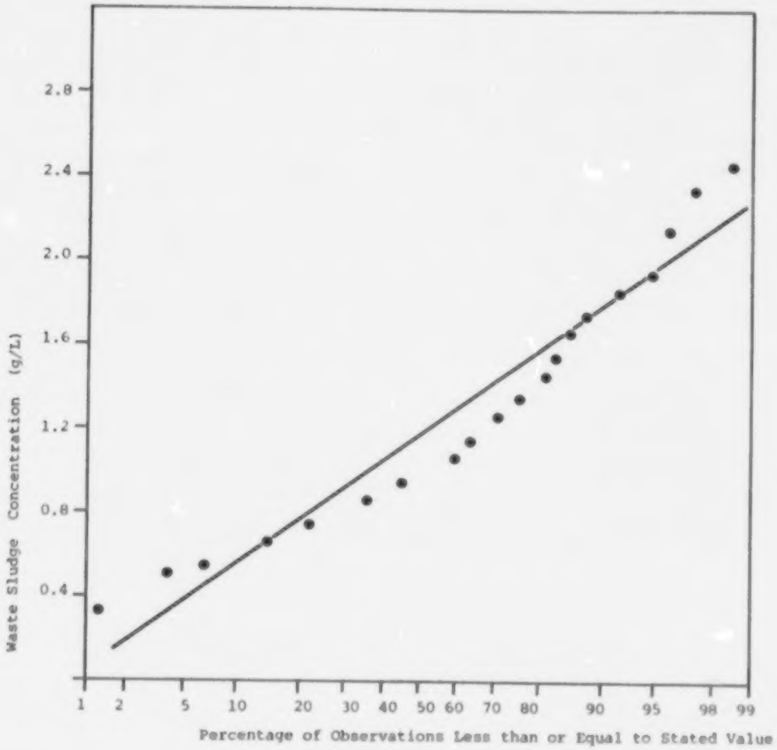


FIGURE 4.13 WASTE SLUDGE CONCENTRATION FROM FLUIDIZED BED

fluidized bed biomass. This is evident from the wide range of oxygen utilization rates -- 0.54 to 1.46 g O_2 per g BOD_5 removed (Table 4.8). However, the mean oxygen utilization rate (0.85 g O_2/g BOD_5 removed) is consistent with the oxygen consumption in suspended growth systems. For typical municipal wastewaters, the rate of oxygen consumption varies from 0.5 to 0.7 g O_2 per g BOD_5 for high rate activated sludge systems to 1.5 g O_2/g BOD_5 for extended aeration systems(46). For conventional air activated sludge processes, the oxygen requirement to effect carbon oxidation is approximately 1.0 g O_2 per g BOD_5 removed(46). Jeris et al(7) reported an oxygen consumption during carbon oxidation of 1.1 g O_2/g BOD_5 at an organic loading of approximately 0.62 g BOD_5/g BVS•d and a solids retention time of 3.5 days.

Sutton et al(25) computed that the oxygenator at Orillia, similar to that in Burlington, was essentially 100 percent efficient in transferring the oxygen supplied to the oxygenator into the reactor feed. Based on the two steady state runs reported(25), the daily oxygen consumption ratio (kg O_2 applied per kg O_2 required where the kg O_2

TABLE 4.8 OXYGEN UTILIZATION FOR CARBON OXIDATION

Reactor Operating Conditions					Dissolved Oxygen (mg/L)		Total BOD_5 (mg/L)		Oxygen Utilization ($g\ O_2/g\ BOD_5$)
HRT (min)	Recycle (Q_R/Q_I)	Flux (m/min)	F/M (g/g•d)	SRT (days)[a]	Influent	Effluent	Influent	Effluent	
20	2.5	0.57	0.55	5.3	22.4	4.8	82	23	1.04
20	2.5	0.57	0.34	4.4	19.6	4.8	67	22	1.15
30	2	0.32	0.42	2.2	16.4	2.0	110	41	0.63
30	2	0.32	0.19	2.2	15.2	8.0	56	16	0.54
30	2	0.32	0.37	4.9	19.6	7.6	84	20	0.56
30	2	0.32	0.20	5.5	17.6	1.6	47	14	1.46
30	2	0.32	0.32	4.1	20.8	8.0	81	21	0.64
30	2	0.32	0.28	7.8	24.0	7.4	78	16	0.80
								Mean	0.85

required was calculated on the assumption of 1.0 kg kg O_2 required per kg of BOD_5 removed plus 4.57 kg O_2 required per kg of nitrate-N produced) ranged from 0.98 to 1.08. Assessment of the transfer efficiency of the oxygenator was outside the scope of the Burlington investigation.

4.2.7 Combined Carbon Removal-Nitrification. None of the pseudo-steady state runs at Burlington resulted in any significant degree of nitrification in conjunction with carbon oxidation. This is not unexpected as the maximum SRT_a attained under the process operating conditions was 6.2 days. The SRT_a was typically in the range of two to four days. In addition, the relatively rapid changes in conditions were not conducive to the establishment of a nitrifying population.

During the Dorr-Oliver studies in Orillia, combined carbon oxidation-nitrification was achieved(25). The Orillia plant was operated longer at a specific condition; and lower organic loadings were applied due to the relatively high suspended solids to BOD_5 ratio in the reactor feed. Based on the Orillia data, it appears that combined carbon removal-nitrification is achievable in the biological fluidized bed.

4.3 Summary

The performance of the process is limited by the system's ability to remove suspended solids, rather than the ability to remove organic carbon from the wastewater. An effluent containing less than 15 mg/L of filterable BOD_5 can be produced consistently at an organic loading approaching 1.0 g total BOD_5/g BVS•d, equivalent to a processing loading of approximately 10 kg total BOD_5/m^3•d. This is about 20 times the typical loading applied to a conventional activated sludge system. However, under these conditions, suspended solids would contribute an additional 20 mg/L BOD_5 to the final effluent. To maintain an effluent containing less than 15 mg/L total BOD_5 requires operation at loadings similar to those applied to a conventional activated sludge system (0.4 kg total BOD_5/m^3•d). Even under these conditions, it is difficult to maintain an effluent suspended solids concentration of less than 15 mg/L with reactor feeds similar to those at Burlington. Due to the impact of effluent suspended solids on fluidized bed performance and design, considerable additional work was conducted to characterize the solids discharged in the treated effluent (Section 7).

5 EFFECT OF DIURNAL LOADING CONDITIONS ON FLUIDIZED BED PERFORMANCE

High-rate, short-retention-time processes tend to be susceptible to changes in influent and are generally less capable of maintaining process performance under variable loading conditions than are systems with a greater hydraulic equalizing capacity. Furthermore, the pseudo-steady state investigations identified influent concentration and process organic and suspended solids loadings as significant variables affecting performance. Therefore, the stability of the fluidized bed system under diurnally varying flow, concentration and loading conditions was evaluated during the carbon oxidation phase.

5.1 Experimental Program

The response of the system under diurnally varying loading conditions was evaluated by imposing a diurnal flow pattern, synchronous with the natural diurnal fluctuation in waste strength, on the process. A peak:average:minimum variation in influent flow (Q_I) of 2:1:0.5 was applied. A real-time digital minicomputer was used to control the feed rate (Q_I), the recycle rate (Q_R) and the rate of oxygen feed.

The system was operated in two modes during the diurnal loading studies to compare constant hydraulic flux conditions and variable hydraulic flux conditions. For the constant conditions, the recycle ratio (Q_R/Q_I) was manipulated in parallel with the imposed diurnal variation in feed rate (Q_I) to maintain a constant total flow ($Q_I + Q_R$) (Figure 5.1). For the variable hydraulic flux, a constant recycle rate (Q_R) was maintained so that the total hydraulic flux on the fluidized bed varied from 0.56 $m^3/m^2 \cdot$min to 0.81 $m^3/m^2 \cdot$min (Figure 5.2). The average operating conditions were identical (Table 5.1).

After comparing performance under constant and variable hydraulic flux conditions, the system was operated for four weeks under variable hydraulic flux conditions.

Monitoring of process performance under diurnal operating conditions included 24-hour flow-proportional composite sampling and hourly discrete sampling to indicate the variability of the effluent quality in response to the diurnal loading.

5.2 Experimental Results

5.2.1 Comparison of Performance under Constant and Variable Hydraulic Flux Conditions.

Each hydraulic flux condition was imposed on the system for 10 days. Twenty-four hour composite samples and discrete samples, taken on one occasion, were used to compare the relative performance of the process in the two operating modes.

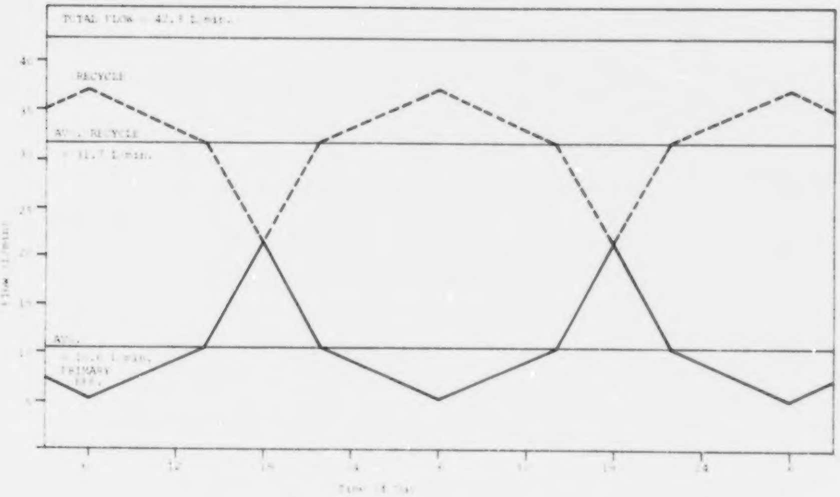


FIGURE 5.1 FLOW PATTERNS: DIURNAL FLOW STUDIES - CONSTANT HYDRAULIC FLUX

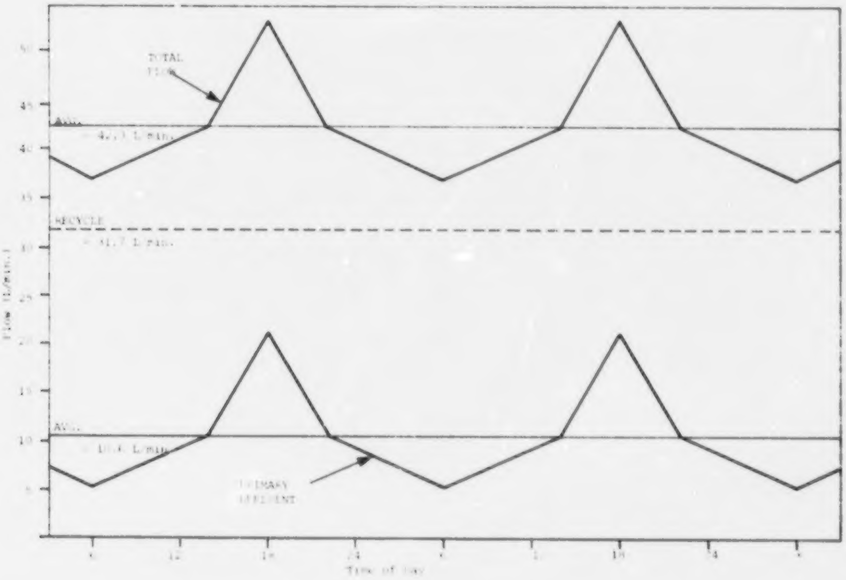


FIGURE 5.2 FLOW PATTERNS: DIURNAL FLOW STUDIES - VARIABLE HYDRAULIC FLUX

TABLE 5.1 AVERAGE OPERATING CONDITIONS FOR DIURNAL LOADING INVESTIGATIONS

	Constant Hydraulic Flux Conditions	Variable Hydraulic Flux Conditions
HRT* (min)		
Maximum	40	40
Minimum	10	10
Average	20	20
Recycle ratio (Q_R/Q_I)		
Maximum	7.0	6.0
Minimum	1.0	1.5
Average	3.0	3.0
Hydraulic flux ($m^3/m^2 \cdot min$)		
Maximum	0.65	0.81
Minimum	0.65	0.56
Average	0.65	0.65
Average effluent		
DO concentration (mg/L)	5	5
Temperature (°C)	16	16

*Hydraulic retention time based on primary effluent feed rate (Q_I) and empty bed reactor volume.

The performance, based on 24-hour composite sampling of influent and effluent streams is compared for the two operating modes in Table 5.2. The characteristics of the fluidized bed during these two 10-day periods were similar. Under constant hydraulic flux conditions, the mean bed volatile solids concentration was 9.61 g/L; during the subsequent variable hydraulic flux period, it was 10.2 g/L.

The raw wastewater during constant hydraulic flux operation was significantly more concentrated, in terms of total and dissolved organics (COD and FOC) and suspended solids, than during the variable period. While the effluent quality during the latter period was slightly better, the difference in FOC and suspended solids was not significant at the

TABLE 5.2 COMPARISON OF PROCESS PERFORMANCE UNDER CONSTANT AND VARIABLE FLUX DIURNAL FLOW CONDITIONS - COMPOSITE SAMPLING RESULTS

		Average Concentration (mg/L)	
	Number of Samples	Constant Hydraulic Flux	Variable Hydraulic Flux
Influent Quality			
Total COD	5	300	248
Filtered COD	5	154	120
Filtered organic carbon	8	32	27
Total BOD_5	1	111	86
Filtered BOD_5	1	66	39
TKN	5	23.7	24.1
Filtered TKN	5	21.1	21.6
Suspended solids	8	91	81
Effluent Quality			
Total COD	5	147	113
Filtered COD	5	101	68
Filtered organic carbon	8	20	18
Total BOD_5	1	41	28
Filtered BOD_5	1	22	11
TKN	5	22.1	22.3
Filtered TKN	5	19.8	20.7
Suspended solids	8	31	28

95 percent confidence level. However, there was no significant difference in removal efficiency.

The hourly variation in effluent quality, in terms of FOC and suspended solids concentrations, in response to the diurnal loading conditions is shown in Figure 5.3 (constant hydraulic flux) and Figure 5.4 (variable hydraulic flux).

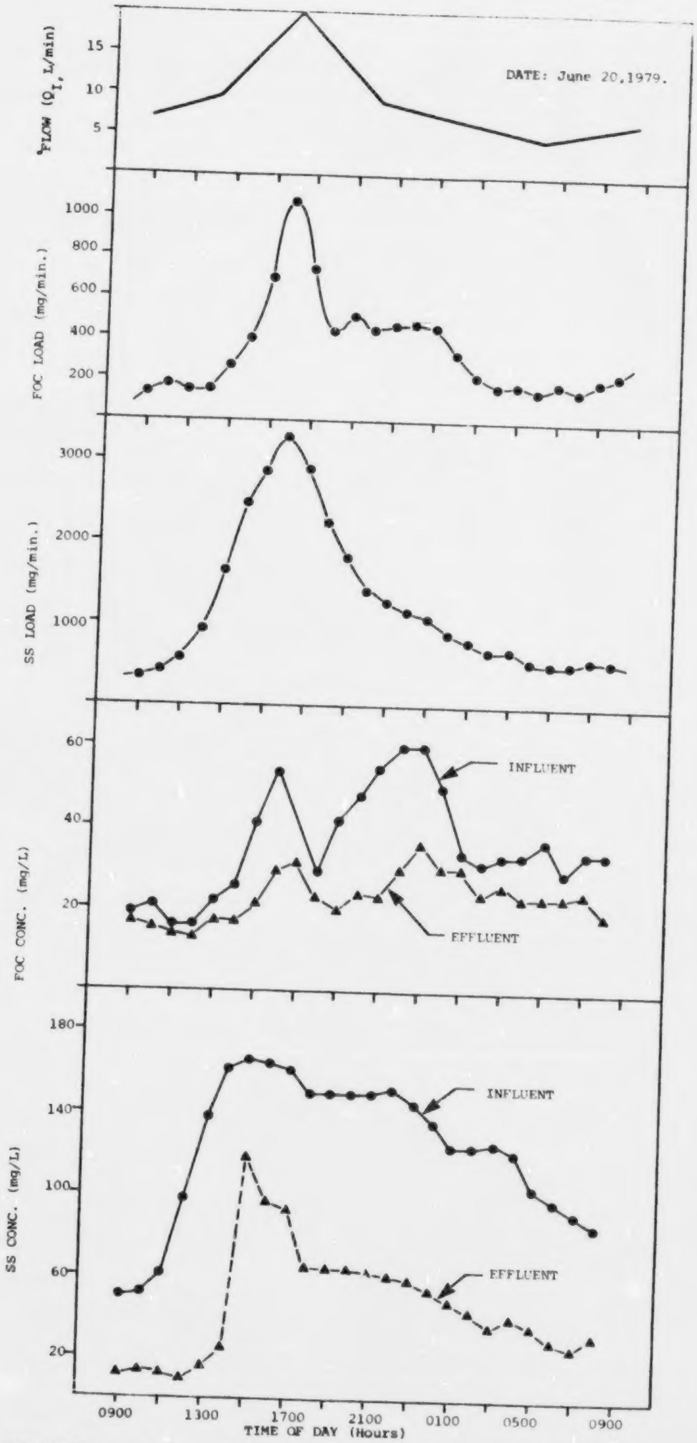


FIGURE 5.3 RESPONSE TO CONSTANT HYDRAULIC FLUX DIURNAL LOADING CONDITIONS

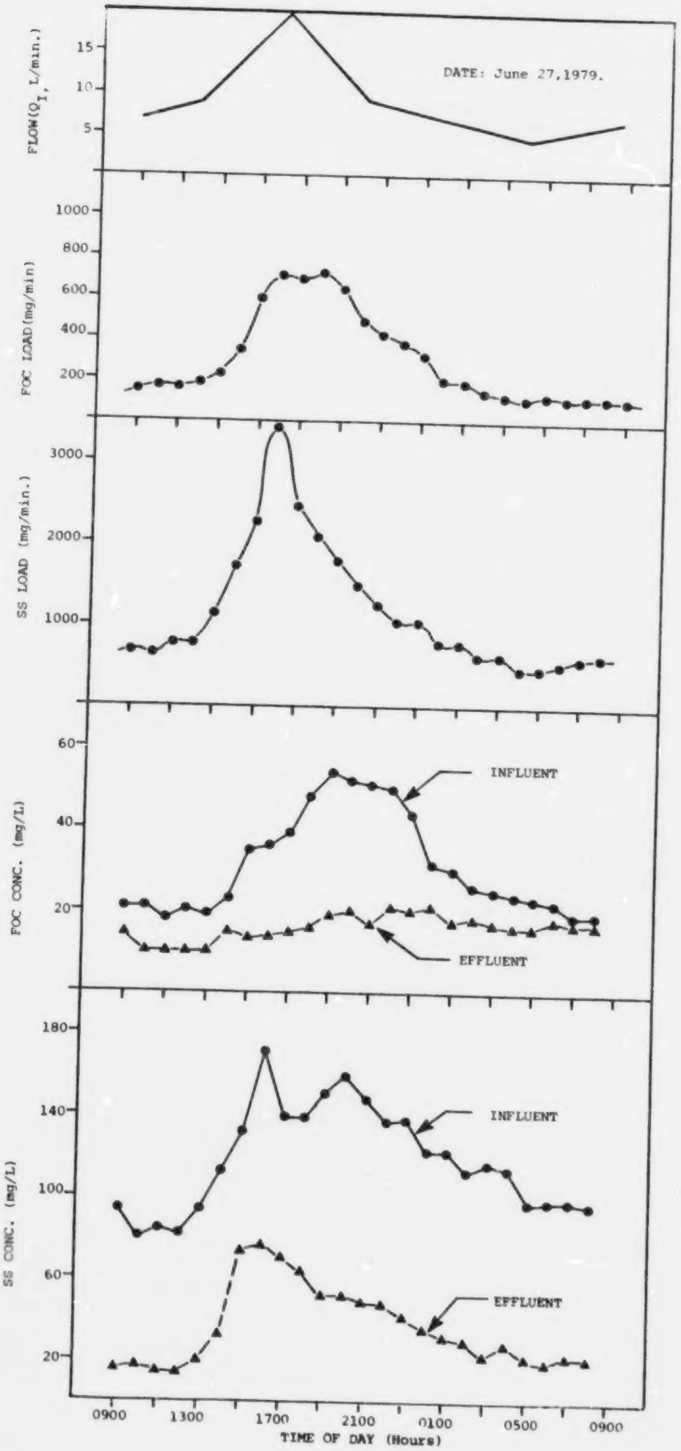


FIGURE 5.4 RESPONSE TO VARIABLE HYDRAULIC FLUX DIURNAL LOADING CONDITIONS

During the discrete sampling at constant hydraulic flux conditions (June 20, 1979), the peak influent concentrations of FOC and suspended solids coincided closely to the peak influent flowrate, and were substantially greater, relative to the mean level, than the 2:1 ratio of peak to average flow imposed on the system.

The peak process loading, in terms of filterable organic carbon, occurred at approximately 1600 hours. The process responded with an immediate increase in effluent FOC. At 2300 hours, there was a second significant increase in effluent FOC concentration coinciding with increased influent FOC. However, the process organic carbon loading was relatively constant during the latter period. The data indicate that the effluent quality in terms of FOC is significantly affected by both the influent wastewater strength and the process loading, as in the steady state investigations.

Peak effluent suspended solids concentration occurred at approximately 1500 hours. However, the magnitude of the response is misleading; the time of peak process loading also coincided with the once-daily return of sand from the sand biomass separation system to the fluidized bed reactor. Therefore, the peak effluent suspended solids concentration measured was a function of both the increased loading and the loss of excess biomass in the effluent from the sand return operation. A smooth curve between the data collected at 1400 hours and 1800 hours would probably more closely approximate the process response. The difference between the interpolated data and the actual data would be a measure of the contribution of the sand return operation to the effluent suspended solids concentration.

During the discrete sampling at variable hydraulic flux conditions (June 27, 1979), the peak influent solids concentration also coincided closely with the peak flowrate imposed on the system. The response, in terms of effluent suspended solids concentration, was similar to that observed for constant hydraulic loading conditions. The impact of the sand return operation, at approximately 1500 hours, obscured the actual impact of the increased process loading on the treated effluent quality.

On June 27, because the peak influent FOC concentration lagged the peak flow by several hours, the increase in organic loading was not as dramatic as observed on June 20. As a result, the increase in effluent organic carbon concentration was not as evident from the June 27 data.

The relative performances of the fluidized bed system, based on the discrete sampling data under constant and variable flow conditions, are compared in Table 5.3. These data appear to indicate that the performance under variable hydraulic loading

TABLE 5.3 COMPARISON OF PROCESS PERFORMANCE UNDER CONSTANT AND VARIABLE FLUX DIURNAL FLOW CONDITIONS - DISCRETE SAMPLING RESULTS

	Constant Hydraulic Flux Conditions	Variable Hydraulic Flux Conditions
Primary Effluent Flow (L/min)		
Maximum	21.2	21.2
Minimum	5.31	5.31
Average	10.6	10.6
FOC Loading (mg/min)		
Maximum	1070	720
Minimum	137	122
Average	345	310
SS Loading (mg/min)		
Maximum	3250	3400
Minimum	360	480
Average	1250	1170
Reactor Feed Quality (mg/L)		
FOC - Maximum	60	54
Minimum	16	18
Average	40	31
SS - Maximum	166	172
Minimum	50	80
Average	124	118
Effluent Quality (mg/L)		
FOC - Maximum	36	21
Minimum	13	10
Average	23	16
SS - Maximum	118	76
Minimum	9	15
Average	46	36

conditions was slightly better than under constant hydraulic loading conditions. However, due to the minimal amount of data available and the differences in the influent characteristics, this cannot be verified statistically. As there was no statistical difference in the performance between the two operating modes based on the 24-hour composite sampling data, the system was maintained in the variable mode for an additional 12 days to collect further data.

5.2.2 Effect of Diurnal Loading on Process Performance. The performance of the fluidized bed process over a three-week period in the variable hydraulic loading mode is shown in Figure 5.5, in terms of filterable organic carbon concentration, and in Figure 5.6, in terms of suspended solids concentration. The operational conditions were identical to those given in Table 5.1.

The average biomass concentration was 10.8 g/L (range 9.46 to 12.04 g/L). The average organic loading on the process was 0.66 kg total BOD_5/kg BVS•d (0.17 kg FOC/kg BVS•d).

Based on the composite sampling approach, the diurnal fluctuations in flow, wastewater strength and process loading did not affect the ability of the fluidized bed to effect carbon oxidation. The results of the diurnal loading investigations are compared with the results of composite sampling under similar pseudo-steady state operation in Table 5.4. The raw wastewater quality in terms of total and filterable BOD_5 and suspended solids concentrations were very similar.

The BVS concentration was slightly higher (11.8 g/L) during the steady state operation. There was no significant difference in effluent quality based on the filterable BOD_5 concentration.

Diurnal operating conditions reduced the system's ability to remove suspended solids. The specific suspended solids loadings were similar (0.57 kg SS/kg BVS•d during diurnal operation, 0.64 kg SS/kg BVS•d during steady state operation); however, the effluent quality under diurnal loading conditions was significantly poorer than under steady state in terms of the effluent suspended solids concentration. This is reflected in the increased total BOD_5 concentration.

The results from the composite sampling are confirmed by hourly discrete sampling data (Figure 5.7) for three days of diurnal operation. The increase in effluent organic carbon concentration in response to the coincidental increase in influent concentration and flow is relatively small compared to the increase in effluent suspended solids concentration. As already discussed, the effluent solids response on June 27 and July 4 are obscured by the effect of the return of sand to the reactor. On July 11, when the sand was returned before the anticipated peak in raw wastewater strength, the effluent response distinctly differentiates between the increase in effluent suspended solids related to the sand return operation (at approximately 1200 hours) and the increase in response to the peak process loading (at approximately 1700 hours).

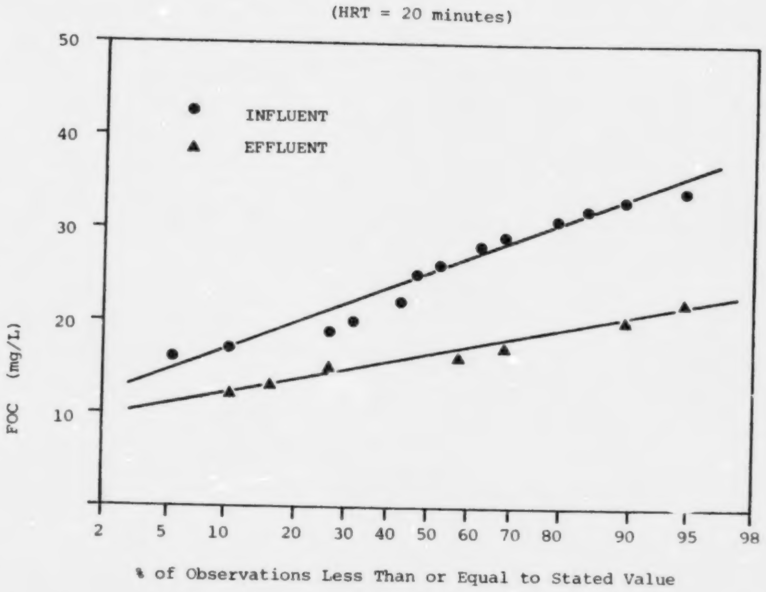


FIGURE 5.5 PROCESS PERFORMANCE UNDER DIURNAL LOADING CONDITIONS: INFLUENT AND EFFLUENT FOC

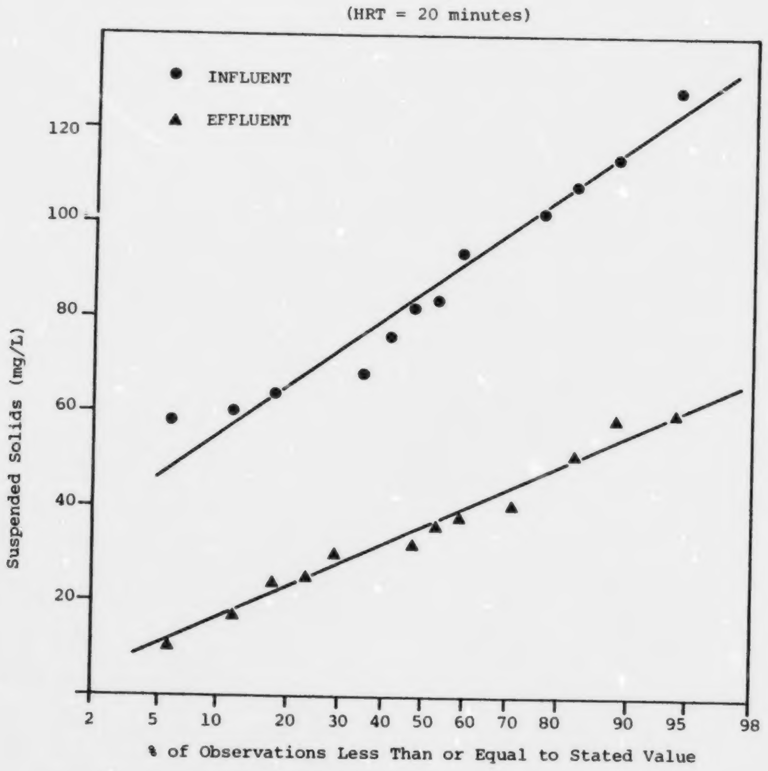


FIGURE 5.6 PROCESS PERFORMANCE UNDER DIURNAL LOADING CONDITIONS: INFLUENT AND EFFLUENT SUSPENDED SOLIDS

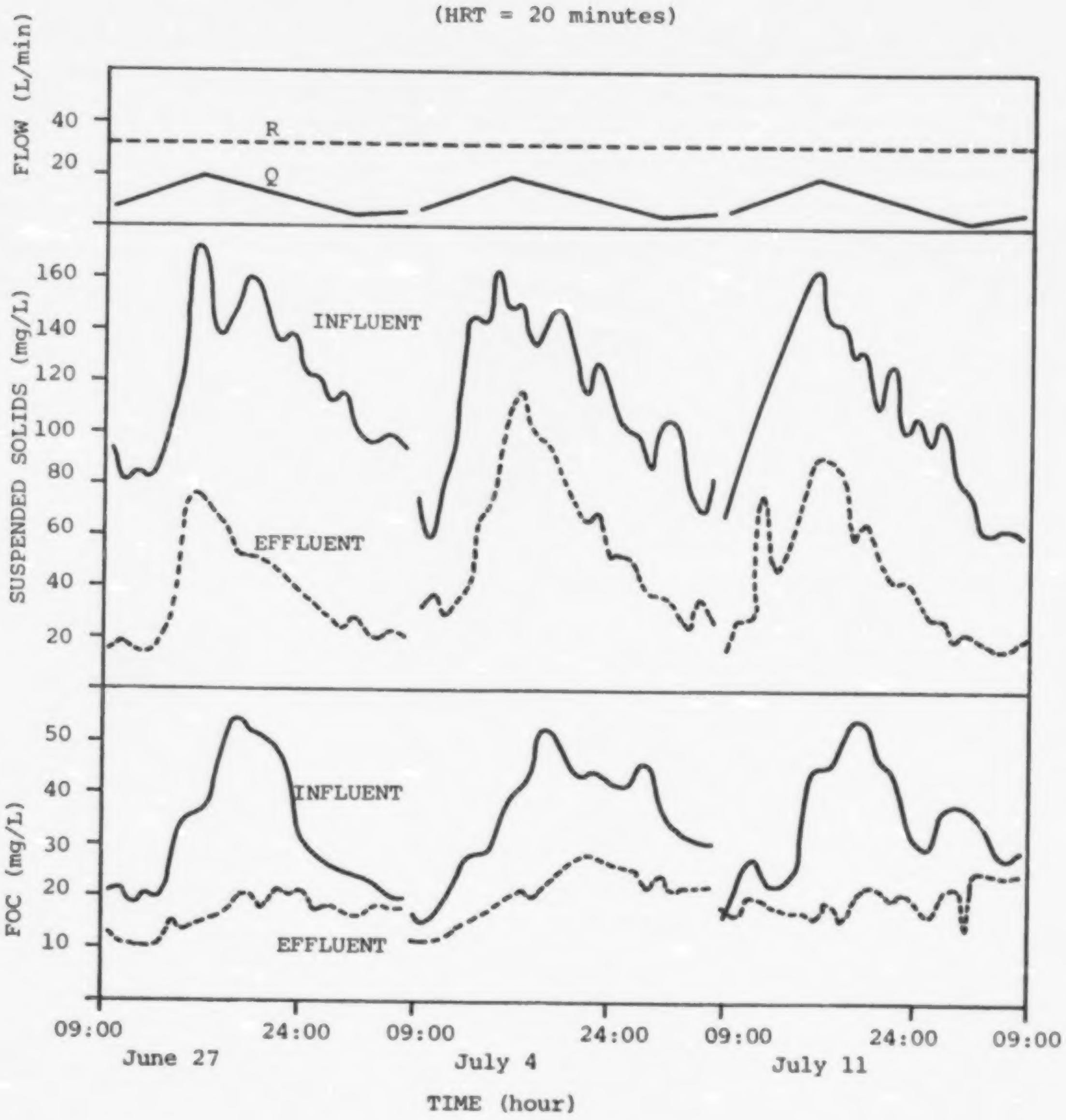


FIGURE 5.7 EFFECT OF DIURNAL LOADING CONDITIONS ON PROCESS PERFORMANCE

TABLE 5.4 COMPARISON OF PSEUDO-STEADY STATE AND DIURNAL OPERATION OF FLUIDIZED BED PROCESS
(HRT = 20 minutes)

Parameter	'Steady State' Operation*		Diurnal Operation**	
	Influent	Effluent	Influent	Effluent
Total BOD_5 (mg/L)	102	28	100	38
Filterable BOD_5 (mg/L)	52	15	46	13
FOC (mg/L)	41	21	25	16
Suspended solids (mg/L)	100	29	86	36

* Steady State operating conditions:

HRT = 20 minutes
Q_R/Q_I = 2.5
T = 12°C
Hydraulic Flux = $0.57\ m^3/m^2\cdot min$

Performance based on average of three consecutive 24-hour composite samples per week for FOC and suspended solids, and one 24-hour composite sample per week for total and filterable BOD_5, over a four-week period.

** Performance based on average of 17 24-hour composite samples for FOC and suspended solids, and three 24-hour composite samples for total and filterable BOD_5.

Due to the relatively high effluent suspended solids concentration produced at a hydraulic retention time of 20 minutes, the system was operated at an HRT of 30 minutes for two weeks to assess the improvement in effluent quality under lower loading conditions. During this period, the system operation was maintained in the variable mode by operating at a constant recycle rate. (See Table 5.5.) The average recycle ratio (Q_I/Q_R) was increased to 5.0 from 3.0 to maintain a total hydraulic flux ($Q_I + Q_R$) similar to that under the earlier diurnal operating conditions.

The performance of the process is shown in Figure 5.8 (influent and effluent FOC) and Figure 5.9 (influent and effluent suspended solids). In Table 5.6, the influent and effluent quality at a 30-minute HRT are compared to similar data at a 20-minute HRT. The raw wastewater in terms of total and filterable BOD_5 under the lower loading conditions (longer HRT) was less concentrated than at the shorter HRT, although there was no significant difference in quality based on FOC and suspended solids.

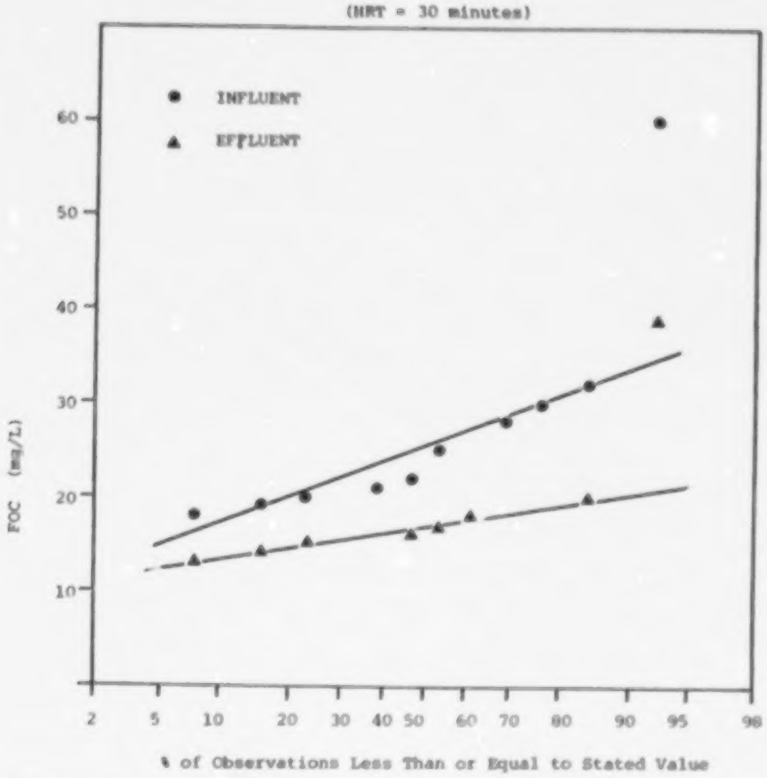


FIGURE 5.8 PROCESS PERFORMANCE UNDER DIURNAL LOADING CONDITIONS: INFLUENT AND EFFLUENT FOC

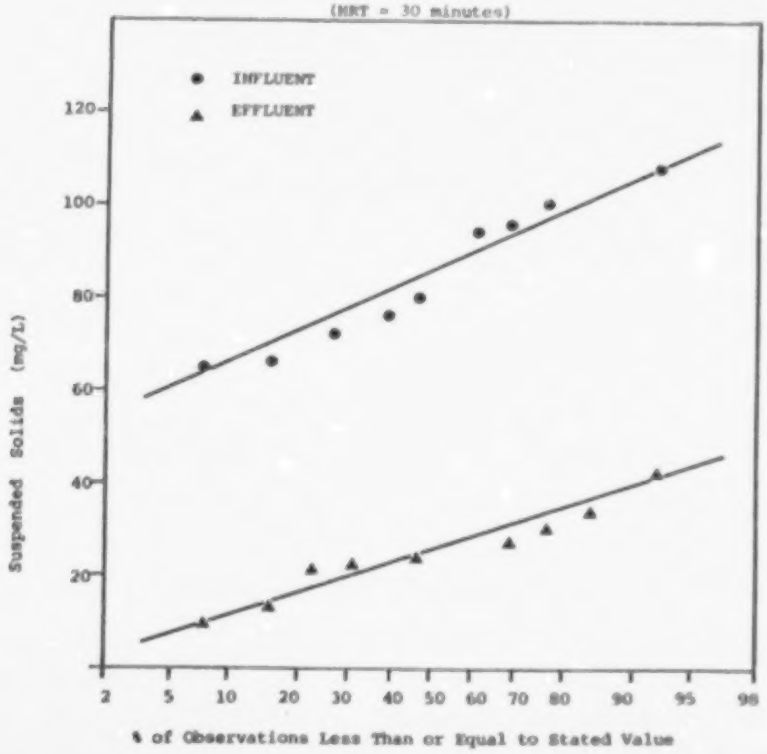


FIGURE 5.9 PROCESS PERFORMANCE UNDER DIURNAL LOADING CONDITIONS: INFLUENT AND EFFLUENT SUSPENDED SOLIDS

TABLE 5.5 AVERAGE OPERATING CONDITIONS FOR DIURNAL LOADING INVESTIGATIONS AT 30-MINUTE HRT

HRT* (min)	
Maximum	60
Minimum	15
Average	30
Recycle Ratio (Q_R/Q_I)	
Maximum	10.0
Minimum	2.5
Average	5.0
Hydraulic Flux ($m^3/m^2\cdot min$)	
Maximum	0.76
Minimum	0.60
Average	0.65

*Hydraulic retention time, based on primary effluent feed rate (Q_I) and empty bed reactor volume.

TABLE 5.6 COMPARISON OF PROCESS PERFORMANCE UNDER DIURNAL LOADING CONDITIONS AT 20 AND 30 MINUTE HRT

Parameter	20-min HRT*		30-min HRT**	
	Influent	Effluent	Influent	Effluent
Total BOD_5 (mg/L)	127	38	82	20
Filterable BOD_5 (mg/L)	63	13	33	7
FOC (mg/L)	25	16	26	17
Suspended solids (mg/L)	86	36	86	26

* For performance basis, see Table 5.4.

** Performance based on average value of 12 24-hour composite samples for FOC and suspended solids and three 24-hour composite samples for total and filterable BOD_5.

The lower process organic loading at the 30-minute HRT (0.19 kg filterable BOD_5/kg BVS•d) compared to the organic loading at the 20-minute HRT (0.42 kg filterable BOD_5/kg BVS•d), along with the lower wastewater strength at the longer HRT, resulted in improved effluent quality in terms of the filterable BOD_5 concentration, as predicted by the pseudo-steady state results. The biomass concentration at the longer HRT conditions was also significantly lower, averaging 8.32 g/L (range 7.30 to 9.48 g/L).

There was a significant improvement in suspended solids removal at the lower loading conditions and an HRT of 30 minutes. The lower suspended solids concentration is partially responsible for the improvement in effluent quality in terms of the effluent total BOD_5. However, as noted in Section 4, attaining a high quality effluent from the process (less than 15 mg/L BOD_5 and suspended solids) is controlled by the system's ability to remove suspended solids.

The performance of the process under diurnal loading conditions at an HRT of 30 minutes is compared with that for similar steady state conditions in Table 5.7. As noted with data collected at an HRT of 20 minutes, diurnal operation had no significant impact on the system's ability to effect carbon oxidation. The effluent suspended solids concentration under diurnal condition was significantly lower (95 percent confidence) than that under steady state conditions, although there was no significant difference in the influent strength. The improvement in effluent quality was minimal.

The results of two days of discrete hourly sampling of the influent and effluent at an HRT of 30 minutes are shown in Figure 5.10. From the FOC response, the diurnal operating conditions had negligible effect on the ability of the process to effect carbon oxidation. Unfortunately, in both cases, the variation in influent strength was small compared to similar data collected at the lower HRT (Figure 5.7) and the peak FOC concentration did not occur at the same time as the peak flow. The maximum variation in effluent FOC was from 9 to 15 mg/L over the two days and probably represents the fraction of non-biodegradable carbon in the raw wastewater.

Similarly, the system response to increasing influent suspended solids concentration and suspended solids loading was less dramatic at an HRT of 30 minutes than at an HRT of 20 minutes. On July 8 and July 24, there was no return of sand to the process during discrete sampling. Therefore, the data (Figure 5.10) indicate only the response to the imposed diurnal conditions.

On July 24, there was a threefold increase in the influent suspended solids concentration and the peak influent strength occurred concurrently with the peak influent

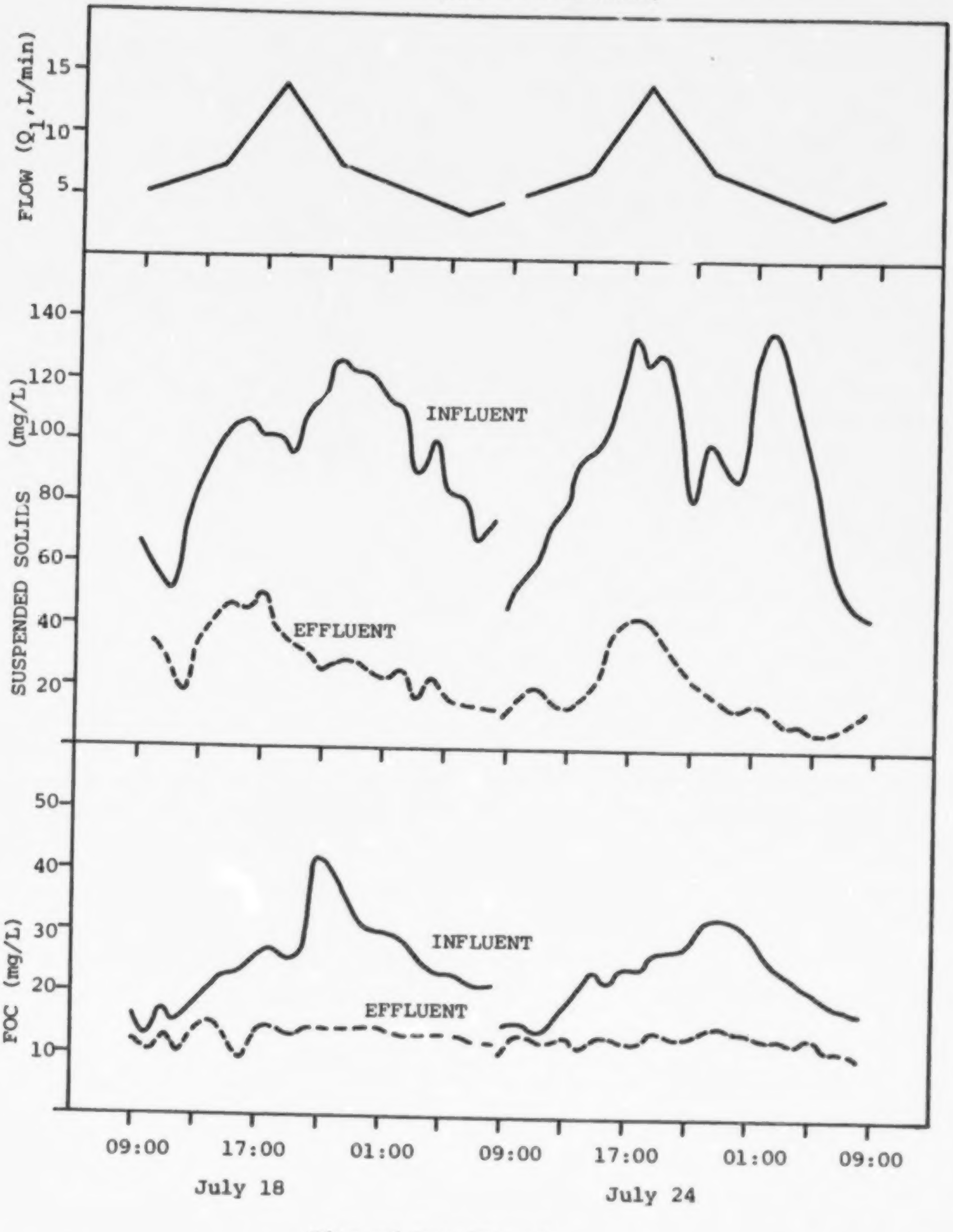

(HRT = 30 minutes)

FIGURE 5.10 EFFECT OF DIURNAL LOADING CONDITIONS ON PROCESS PERFORMANCE

TABLE 5.7 COMPARISON OF PSEUDO-STEADY STATE AND DIURNAL OPERATION OF FLUIDIZED BED PROCESS
(HRT = 30 minutes)

Parameter	Steady State Operation*		Diurnal Operation**	
	Influent	Effluent	Influent	Effluent
Total BOD_5 (mg/L)	83	21	82	20
Filterable BOD_5 (mg/L)	34	9	33	7
FOC (mg/L)	29	18	26	17
Suspended solids (mg/L)	97	35	86	26

* Steady State operating conditions:
HRT = 30 minutes
Q_R/Q_I = 3
T = 8 - 16°C
Hydraulic Flux = $0.43\ m^3/m^2 \cdot min$

Performance based on average of three consecutive 24-hour composite samples per week for FOC and suspended solids and one 24-hour composite sample per week for total and filterable BOD_5 over a three-week period.

** For performance basis, see Table 5.6.

flow (Q_I). The resultant eightfold increase in process suspended solids loading produced a significant deterioration in effluent quality; however, the process gain was considerably less than observed at the lower HRT.

5.3 Variability of Process Performance

The six weeks of operation under diurnally varying loading conditions were used to assess the variability of effluent quality produced. The process performance is summarized in Table 5.8 and shown graphically in Figure 5.11 (FOC and suspended solids) and Figure 5.12 (COD and BOD_5). The system was actually operated under three different conditions; constant and variable flux at 20-minute HRT, and variable flux at 30-minute HRT. Therefore, the variation is greater than could be expected over a similar length of time at a fixed operating condition.

The effluent quality in terms of filterable organic carbon concentration was relatively consistent except for an extremely high peak concentration (32 mg/L on Day 35) in response to a high influent FOC concentration (60 mg/L). The other influent

TABLE 5.8 FLUIDIZED BED PROCESS PERFORMANCE DURING DIURNAL OPERATION

Date*	Influent Quality (mg/L)						Effluent Quality (mg/L)						
	COD		BOD_5**				COD		BOD_5**				
	Total	Filtered	Total	Filtered	FOC	SS	Total	Filtered	Total	Filtered	FOC	SS	Bed Volatile Solids (g/L)
June 13/79	309	160	124	77	46	102	160	98	54	34	21	37	8.43
June 14	298	137	119	61	25	110	128	83	39	23	17	38	
June 15					36	75					16	12	
June 16					22	62					16	13	10.05
June 17					21	84					18	34	
June 18	298	169	119	83	32	80	151	113	50	45	19	33	10.34
June 19	278	144	111+	66+	31	104	151	103	41+	22+	19	35	
June 20	308	165	123	80	36	124	157	107	53	40	23	46	
June 21	319	160	129	77	46	112	146	107	48	40	30	18	
June 22					32	68					20	32	
June 23					19	76					16	10	
June 24	242	114	93	45	19	60	122	84	37	25	16	30	
June 25	282	139	112	62	34	114	125	84	38	25	22	38	10.29
June 26	275	125	108	53	33	102	111	53	32	4	20	32	
June 27	270	131	106	57	31	118	124	78	39	22	16	36	10.13
June 28	230	108	86+	39+	28	102	119	63	28+	11+	17	40	
June 29	209	114	77	45	31	68	87	56	24	6	16	24	10.29
June 30					19	58					15	17	
July 1					16	68					12	25	
July 2					20						13		
July 3	259	100	93+	47+	26	82	140	71	38+	12+	18	51	9.46
July 4	277	130	109	56	35	112	145	81	47	23	20	57	11.21
July 5	268	120	105	49	28	108	131	66	41	13	16	51	
July 6					29	84					18	60	11.32
July 7					22	82					16	40	
July 8					17	64					12	32	
July 9	273	145	107	66	31	102	137	74	43	18	17	36	11.52
July 10	248	91	96	30	22		123	58	37	7	15		
July 11	292	132	116	57	34	108	142	89	46	28	19	44	12.04
July 12	275	108	120+	51+	25	128	141	71	50+	15+	16	59	
July 13					28	108					20	40	9.48
July 14					21	76					16	36	
July 15					21	66					16	38	
July 16	199	89	72	28	19	94	93	49	24	2	13	41	7.74
July 17	259	116	101	47	60	80	97	65	25	12	39	24	
July 18	258	108	100	41	25	94	106	58	29	8	13	23	8.63
July 19	203	86	89+	36+	22	94	87	56	22+	6+	17	27	
July 20	265	108	103	41	32	76	70	53	15	4	16	13	7.30
July 21					18	65					15	9	
July 22					20	72					14	22	
July 23	187	103	67	38	30	100	110	52	31	3	18	30	7.79
July 24	225	123	108+	51	23	94	103	65	20+	12	13	37	
July 25	226	99	85	35	28	108	97	52	25	3	20	34	8.95
July 26	236	100	75+	30+	25	96	84	59	18+	7+	20	21	

Notes:

*Operating Conditions:

	June 13 - June 21 -	Constant Hydraulic Flux, HRT = 20 minutes
	June 22 - July 12 -	Variable Hydraulic Flux, HRT = 20 minutes
	July 13 - July 26 -	Variable Hydraulic Flux, HRT = 30 minutes

** All BOD_5 results marked are analytical results. Other BOD_5 values are calculated based on the following correlations with the COD values:

Total BOD_5 = 0.467 x Total COD - 21
Filtered BOD_5 = 0.683 x Filtered COD - 32

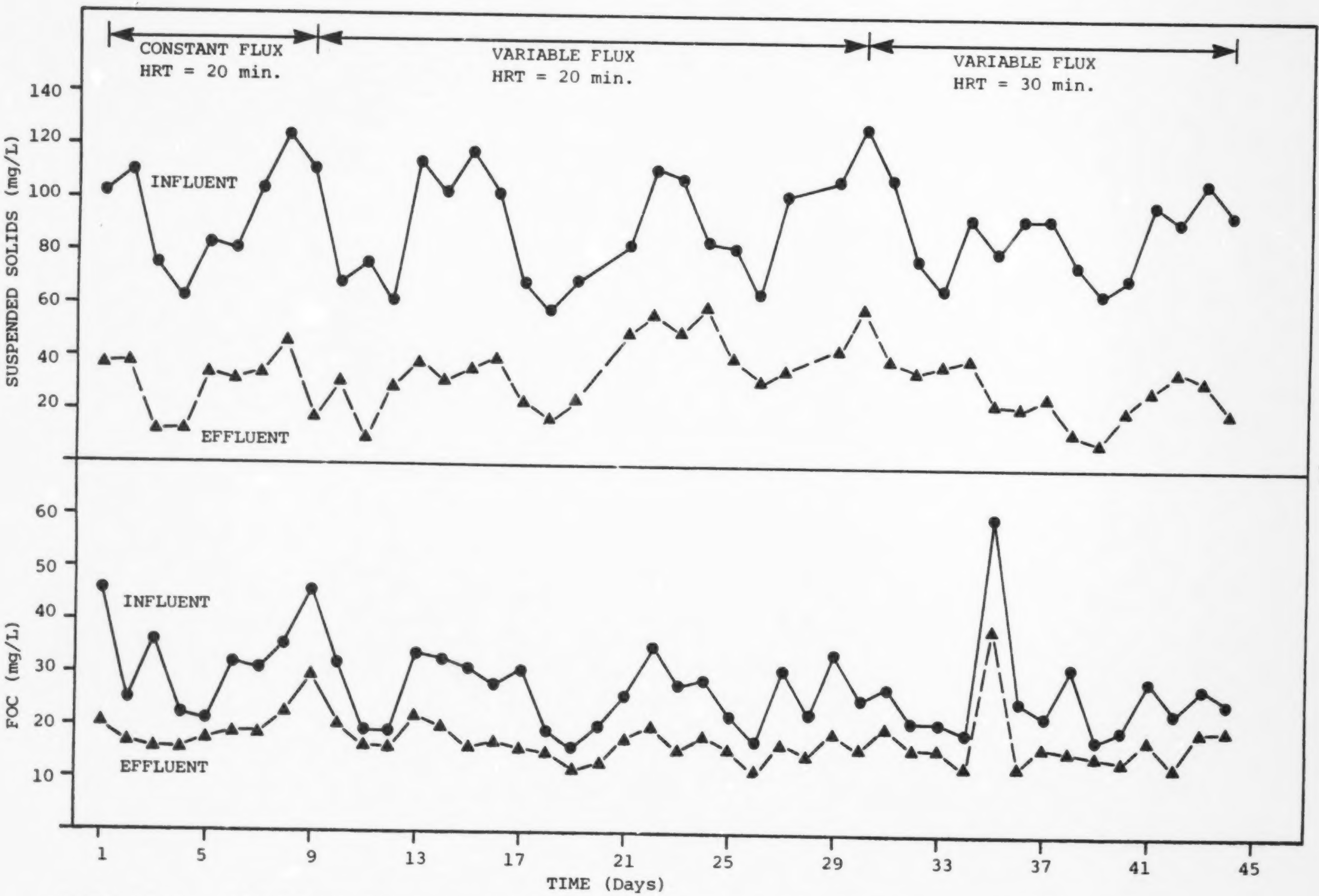


FIGURE 5.11 DAILY VARIATION IN FLUIDIZED BED PERFORMANCE - FOC AND SUSPENDED SOLIDS

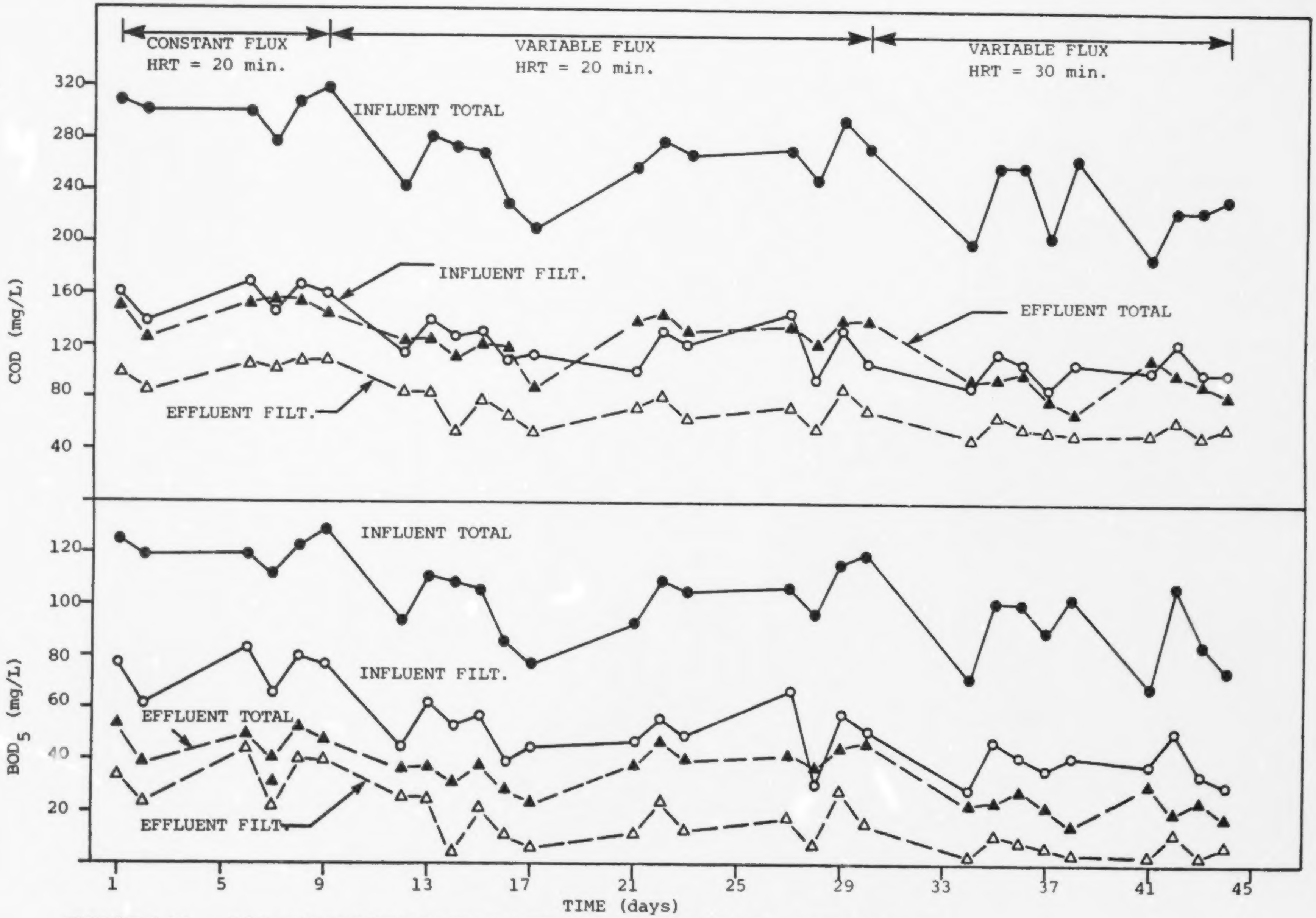


FIGURE 5.12 DAILY VARIATION IN FLUIDIZED BED PERFORMANCE - COD AND BOD_5

and effluent quality parameters (total and filterable COD, suspended solids) did not reflect this. It was concluded that these FOC data were in error and they are not included in the subsequent analysis.

Effluent quality in terms of suspended solids concentration showed more daily variation than similar data for FOC. This was also reflected in the total COD and total BOD_5 data compared to the filterable COD and filterable BOD_5 data shown in Figure 5.12. The fluidized bed process can maintain consistent performance over a range of loadings in terms of organic carbon removal; however, the inability to maintain a high degree of suspended solids removal is reflected by a wide variation in effluent quality in terms of suspended solids and total BOD_5 concentrations.

Based on the six weeks of operating data (see Table 5.9) the average effluent quality was 33 mg/L suspended solids and 36 mg/L total BOD_5. However, the expected range of effluent quality (95 percent confidence) was 8 to 58 mg/L suspended solids and 14 to 58 mg/L total BOD_5. Analysis of the experimental data at 30-minute HRT (Table 5.10) indicates an improvement in effluent quality to 28 mg/L suspended solids and 23 mg/L total BOD_5, and a reduction in the degree of variability of the effluent quality. Effluent suspended solids concentration would be expected to vary from 8 to 48 mg/L.

5.4 Solids Yield

The diurnal operating data allow a better estimate of the net solids yield from the fluidized bed process than do the pseudo-steady state data because the system was operated under constant conditions for a longer period during the diurnal studies.

As discussed in Section 4.2.5, the calculated net yield coefficient and solids retention time (SRT) will depend on whether the effluent suspended solids are considered to be excess biomass generated from biological oxidation in the fluidized bed reactor. In a suspended growth biological treatment system, the conventional SRT is calculated according to the following equation:

$$SRT_c = \frac{\text{volatile solids in the bioreactor}}{\text{volatile solids lost in effluent or intentionally wasted}} \qquad (5.1)$$

In such systems, the effluent solids are considered to be excess biomass and are included in the total mass of biological solids lost from the process.

TABLE 5.9 VARIATION IN EFFLUENT QUALITY - GROUPED DATA

	Effluent Quality (mg/L)			
Parameter	Mean	Standard Deviation	Coefficient of Variation (%)	95% Confidence Region*
Total COD	121	24.7	20.4	75-169
Filterable COD	73	19.1	26.2	36-110
Total BOD_5	36	11.2	31.1	14-58
Filterable BOD_5	17	12.3	72.4	0-41
FOC	18	4.6	25.6	9-27
Suspended Solids	33	12.8	38.8	8-58

*Assuming parameter is normally distributed.

TABLE 5.10 VARIATION IN EFFLUENT QUALITY - 30-MINUTE HRT

	Effluent Quality (mg/L)			
Parameter	Mean	Standard Deviation	Coefficient of Variation (%)	95% Confidence Region*
Total COD	94	12.4	13.2	70-118
Filterable COD	57	5.7	10.0	46-68
Total BOD_5	23	5.1	22.2	13-33
Filterable BOD_5	6	3.8	63.3	0-13
FOC	16	2.7	16.9	11-21
Suspended Solids	28	10.0	35.7	8-48

*Assuming parameter is normally distributed.

In the fluidized bed process, the effluent solids constitute, to a large extent, influent solids not removed by the reactor. This was demonstrated by Sutton et al(9) and verified during this study, based on ATP comparisons of the influent and effluent solids and the reactor biomass (see Section 7.1). On this basis, a modified SRT can be calculated according to the following equation:

$$SRT_m = \frac{\text{volatile solids in the bioreactor}}{\text{volatile solids intentionally wasted}} \tag{5.2}$$

The actual process SRT, SRT_a, will be somewhere between SRT_c and SRT_m as a fraction of the effluent solids do represent biomass generated in the biological reactor. Based on the ATP measurements (Section 7.1), the effluent solids consist of approximately 65 percent uncaptured influent solids and 35 percent biomass sloughed from the reactor support media. The actual process SRT can be estimated according to the following equation:

$$SRT_a = \frac{\text{volatile solids in the reactor}}{(\text{volatile solids intentionally wasted}) + 0.35\,(\text{volatile solids in effluent})} \tag{5.3}$$

The yield coefficient, Y, and the process SRT are interrelated according to the equation:

$$\frac{F}{M} = \frac{1}{SRT} \cdot \frac{1}{Y} \cdot \frac{1}{E} \tag{5.4}$$

where E is the fraction of BOD_5 removed:

$$E = \frac{\text{Influent } BOD_5 - \text{Effluent } BOD_5}{\text{Influent } BOD_5} \tag{5.5}$$

For the calculation of the Y_c, corresponding to the conventional SRT_c, E was based on the influent total BOD_5 and the effluent filtered BOD_5 according to the convention adopted by the Water Pollution Control Federation and the American Society of Civil Engineers(40). However, for the calculation of Y_m and Y_a, corresponding to SRT_m and SRT_a, the BOD_5 removal efficiency was based on the removal of total BOD_5 across the reactor, as the effluent suspended solids are not considered to be biomass.

The process SRT and yield coefficients calculated by each method for the diurnal experimentation at 20-minute and 30-minute HRT are summarized in Table 5.11. The values SRT_a and Y_a provide the best estimate of the system SRT and the process

TABLE 5.11 NET SOLIDS YIELD FROM FLUIDIZED BED SYSTEM

Source	Operating Conditions				Solids Retention Time (days)			Net Yield (mg VS/mg BOD_5)		
	Temp. (°C)	HRT (minutes)	BVS (g/L)	F/M (g/g•d)	SRT_c	SRT_a	SRT_m	Y_c	Y_a	Y_m
This study	16	20	10.46	0.74	2.3	3.4	4.5	0.74	0.65	0.48
	16	30	8.32	0.52	3.6	5.0	6.3	0.58	0.52	0.41
Sutton *et al*(25)	11	26	15.08	0.29	3.8	5.1	6.5	0.99	0.85	0.76
	11	36	11.69	0.18	5.1	6.8	9.5	1.15	0.99	0.82
	21	37	14.56	0.17	8.5	12.0	19.5	0.76	0.61	0.43

yield coefficient. Also included in Table 5.11 are similar data from steady state experimentation at Orillia, Ontario(25). In this case, Y_c-SRT_c and Y_m-SRT_m were calculated according to the same approach used in this study. However, estimated Y_a-SRT_a values were based on the assumption that the minimum effluent suspended solids measured under specific operating conditions represented the continuous background contribution of biomass from the reactor to the treated effluent.

The yield coefficient Y_a based on the diurnal data is typical of those reported for conventional suspended growth systems treating municipal wastewaters. Yield coefficients for such systems are generally reported in the range 0.50(41) to 0.70 mg VS/mg BOD_5(40). The yields from the Burlington pilot plant were slightly lower than those reported by Dorr-Oliver(25). Sutton et al(25) claim that the higher yields in Orillia result from the abnormally high suspended solids to BOD_5 ratio in that particular wastewater.

5.5 Summary

Operation of the system under diurnally varying flow (peak: average: minimum variation = 2:1:0.5) and waste strength had no significant impact on carbon oxidation. Although discrete hourly sampling indicated a measurable response in terms of effluent organic carbon concentration to the peak organic loading, the average effluent quality based on 24-hour composite sampling was not significantly different from that attained under similar steady state conditions. It appears that the fluidized bed system is adequate to withstand relatively large increases in organic carbon loading.

Increasing waste strength and process suspended solids loading, resulting from diurnal operating conditions, produced a rapid deterioration in effluent quality in terms of suspended solids concentration. The magnitude of the response was more significant at lower system hydraulic retention times. Despite the instantaneous deterioration in effluent quality, the average quality based on 24-hour composite samples collected at an average HRT of 30 minutes was not significantly poorer than at similar steady state operating conditions.

Operation at a variable flux did not affect performance. This simplifies the design of the fluidized bed system. Recycle pumping can be maintained at a constant rate rather than varying continuously in response to increasing raw waste flows to maintain a constant hydraulic flux.

The effluent quality from the fluidized bed reactor varied considerably day to day in terms of suspended solids concentration. This variation was also reflected in the effluent total BOD_5 concentration. Carbon oxidation performance was relatively stable.

The solids yield is comparable to that expected in a conventional activated sludge system treating primary clarified municipal wastewater, in the range 0.5 to 0.7 g VS per g BOD_5 removed.

6 PERFORMANCE OF FLUIDIZED BED UNDER DYNAMIC OPERATING CONDITIONS

The technique of time series analysis, as developed by Box and Jenkins(37), was used to define more precisely the response of the fluidized bed system under dynamic operating conditions, and to identify the variables or combination of variables which most significantly affect the design and operation of the process.

The time series methodology involves monitoring the process response to a predesigned and regulated sequence of perturbations of the influent variables. The resultant models, which relate effluent response to variations in the influent variables, define the most significant variables affecting performance and are a basis for process assessment and possibly control. Using few parameters, these models can closely approximate a system whose true nature is likely much more complex(37). From these dynamic model paramters (δ, ω, b), it is possible to calculate indicators of process response such as the process time constant, τ; the process steady state gain, g (% change in output/% change in input); and the process dead time, which have exact analogues in mechanistic systems(37). Using such characteristic values, one may predict the time to approach a new steady state value after a step change in process input or to return to steady state after a shock load. This is valuable in the descriptive analysis of processes susceptible to inputs variable in time and in the design of schemes to, say, reduce the impact of variable or shock loads or to manipulate other variables for control purposes. In the past, these techniques have been successfully applied to suspended growth(38) and fixed film(39) processes for their descriptive analysis. The technique has been described elsewhere(34).

6.1 Experimental Design

The objectives were to identify causal relationships between:

1) effluent filterable organic carbon concentration (FOC_E) and reactor feed rate (Q_I), influent filterable organic carbon concentration (FOC_I), and process filterable organic carbon loading (FOC_L)

2) effluent suspended solids concentration (SS_E) and reactor feed rate (Q_I), influent suspended solids concentration (SS_I), process suspended solids loading (SS_L) and recycle ratio (Q_R/Q_I).

To allow precise estimation of model parameters and to discriminate between competing models, an input perturbation sequence was designed which eliminated the high correlation between reactor feed rate (Q_I), and influent FOC and suspended solids concentrations normally encountered in municipal wastewater. The reactor feed strength was manipulated by diluting the primary clarifier effluent with tap water. A random (white noise) sequence was applied to the dilution water flow (Q_D), giving random variation in influent concentrations of FOC_I and SS_I. A schematic is shown in Figure 6.1.

To minimize the possible adverse effects of large changes in flow, the input flow (Q_I) sequence after dilution was designed to be auto-correlated and to vary more slowly than the random variations in concentration. Upper and lower limits were placed on the dilution flow (Q_D) and the feed flow (Q_I) so that there was always organic carbon present in the feed to the reactor.

According to the experimental design, the influent variables to the system were manipulated every eight minutes for 16 hours. Seven minutes after each step change in influent variables, the influent and effluent streams were sampled and analyzed for filterable organic carbon and suspended solids. Every twelfth sample was analyzed for total and filtered BOD_5. This interval was selected based on the previously measured response time of the process to a step change in influent FOC concentration. The real-time digital minicomputer was used to monitor and precisely control the influent, dilution and recycle pumps at the pilot plant and to log the data from each of three magnetic flow-meters. Reactor effluent DO was also controlled by the minicomputer.

The sewage stored for the experimentation had a filterable organic carbon concentration of approximately 46 mg/L. To ensure measurable response, 40 mg/L of organic carbon was added to the system feed in the form of dextrose. The bed volatile solids concentration was 9.12 g/L.

The mean operating conditions for the dynamic modelling of the fluidized bed were identical to those in the constant flux diurnal loading investigations (Section 5). The oxygenator and biological reactor system were operated at a constant total hydraulic flux ($Q_I + Q_R$) by simultaneously varying the recycle flow (Q_R) in response to designed perturbations in the reactor feed flow (Q_I). The mean operating conditions for the system were:

Average HRT (empty bed, based on Q_I) = 20 minutes
Total hydraulic flux (constant) = 0.65 $m^3/m^2 \cdot min$

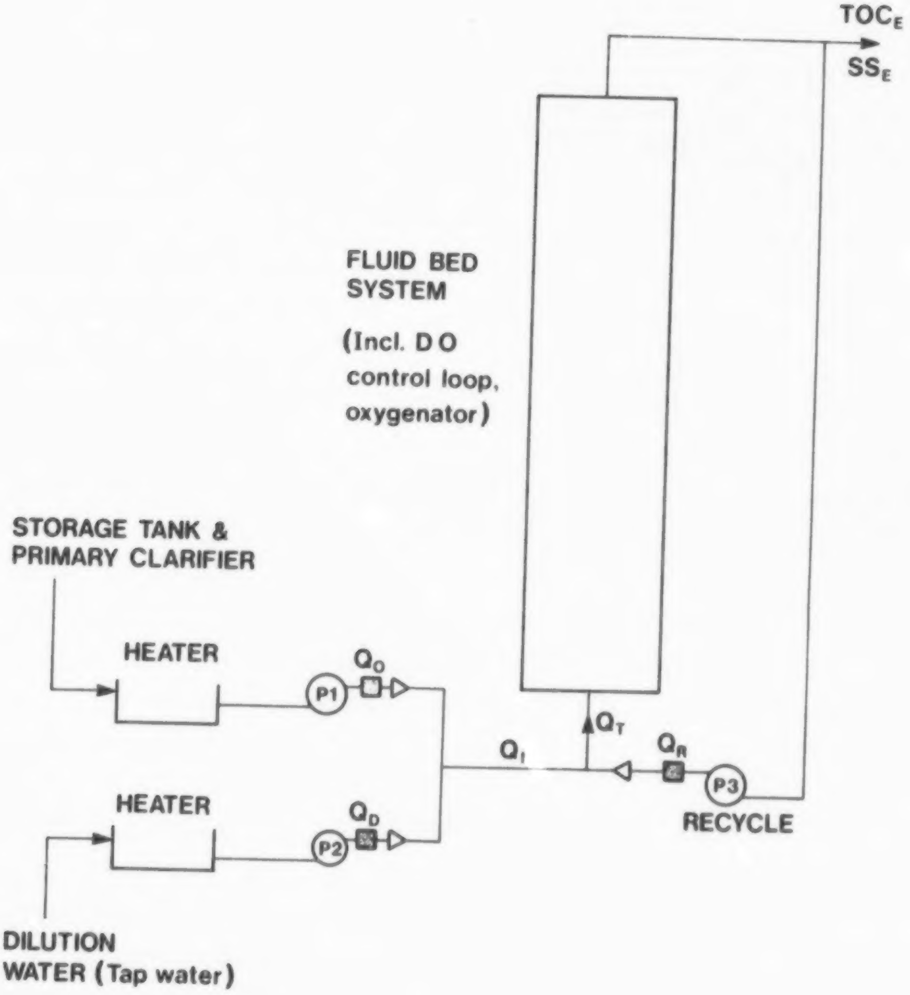

TOC$_E$
SS$_E$
FLUID BED SYSTEM
(Incl. D O control loop, oxygenator)
STORAGE TANK & PRIMARY CLARIFIER
HEATER
P1
Q$_O$
Q$_I$
Q$_T$
Q$_R$
P3
RECYCLE
HEATER
P2
Q$_D$
DILUTION WATER (Tap water)

FIGURE 6.1 EXPERIMENTAL APPARATUS

Average recycle ratio (Q_R/Q_I) = 3.0
Average effluent DO concentration = 5 mg/L
Temperature = 16°C

6.2 Experimental Results

The input FOC concentration and flow (Q_I) sequences and the process response in terms of the effluent FOC concentration are shown in Figure 6.2. Similar data for suspended solids are shown in Figure 6.3. Real-time computer control of the various feed pumps resulted in excellent agreement between the designed and actual input flows, with individual pumps responding rapidly and precisely to each step change in flow (Figure 6.4). Review of the actual influent flow (Q_I) data indicated that it was auto correlated as designed. There was no significant correlation between the influent concentration (FOC_I, SS_I) and the influent flow. The random nature of the actual influent concentration sequences were verified, based on the experimental data.

Based on the measured response of the fluidized bed process, an iterative approach was used to develop discrete dynamic transfer function-noise models of the form:

$$Y_t = \frac{(\omega_0 - \omega_1 B - \omega_2 B^2 - \cdots - \omega_s B^s)B^b}{(1 - \delta_1 B - \delta_2 B^2 - \cdots - \delta_r B^r)} X_t + \frac{(1 - \theta_1 B - \cdots - \theta_q B^q)}{(1-B)^d (1 - \phi_1 B - \cdots - \phi_p B^p)} a_t \quad (6.1)$$

where: Y_t = system response (deviation from the mean) at time t,
X_t = system input (deviation from the mean) at time t,
δ, ω = transfer function model parameters,
r,s = transfer function model orders,
b = system delay period,
B = backward shift operator, e.g. $B^n Y_t = Y_{t-n}$
θ, ϕ = noise model parameters,
p,d,q = noise model orders,
a_t = white noise sequence with zero mean and constant variance.

The forms of the transfer function models were identified based on the cross-correlation between the system input variables and output responses. The forms of the noise models were identified from the residuals of the fitted transfer function models.

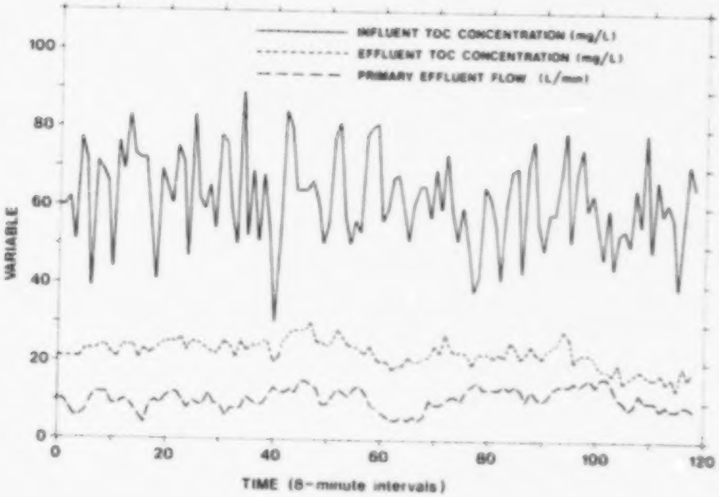


FIGURE 6.2 RESPONSE OF EFFLUENT TOC CONCENTRATION

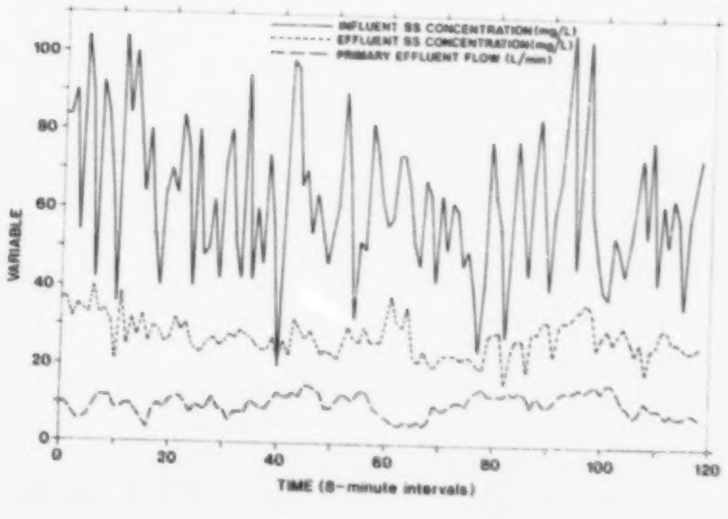


FIGURE 6.3 RESPONSE OF EFFLUENT SUSPENDED SOLIDS CONCENTRATION

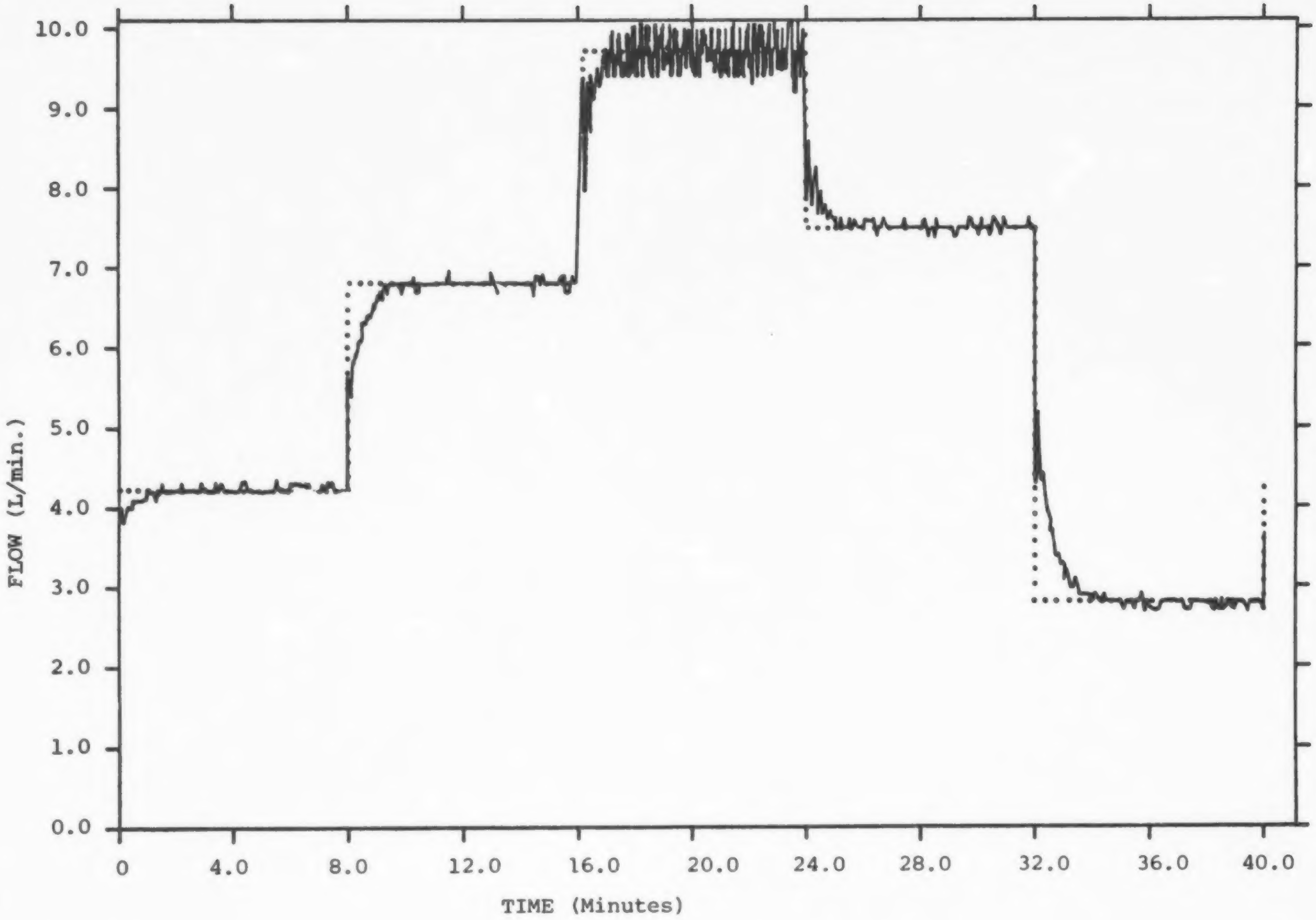


FIGURE 6.4 TYPICAL FLOW RESPONSE DURING TIME SERIES STUDIES

The parameters of the combined transfer function-noise models were estimated using non-linear least squares and the model adequacies were verified by diagnostic checks on the final residuals.

6.3 Cross-Correlation Functions

The cross-correlation function between influent variables and effluent responses indicates whether a given output response is statistically related to changes in a specific input variable. A significant correlation between an input variable X and an output response Y generally indicates a cause-and-effect relationship if a proper statistical procedure has been used$^{(37)}$.

The cross-correlations between effluent FOC and the influent variables (FOC_I, FOC_L, Q_I) are shown in Figure 6.5. The data indicate that, at the 95 percent confidence level, the effluent FOC response is significantly correlated to the influent FOC concentration (FOC_I), the reactor feed rate (Q_I) and the process organic carbon loading (FOC_L). However, the correlation between flow and effluent FOC is relatively weak.

Similar cross-correlation functions relating the effluent suspended solids response to the input variables are shown in Figure 6.6. Based on these, the effluent suspended solids (SS_E) is significantly affected by the influent suspended solids concentration (SS_I) and the process suspended solids loading (SS_L). There was, however, no significant correlation between the effluent suspended solids concentration and the reactor feed rate (Q_I) or the recycle ratio (Q_R/Q_I). The correlation between effluent suspended solids concentration and process suspended solids loading is relatively weak.

It appears that effluent quality in terms of organic carbon concentrations is most significantly affected by the process organic carbon loading and the influent organic carbon concentration. The reactor feed rate has a lesser impact.

In terms of the effluent suspended solids concentration, the influent suspended solids concentration has the most significant impact. However, the process loading (SS_L) also influences the system response. A summary of results is given in Table 6.1.

The organic carbon concentrations are highly correlated with the total and filtered BOD_5 as would be expected since the system feed source was stored sewage, spiked with dextrose and diluted (Figure 6.7). The cross-correlations noted between the input variables and the system response in terms of FOC_E would be equally applicable in terms of total and filtered BOD_5 response.

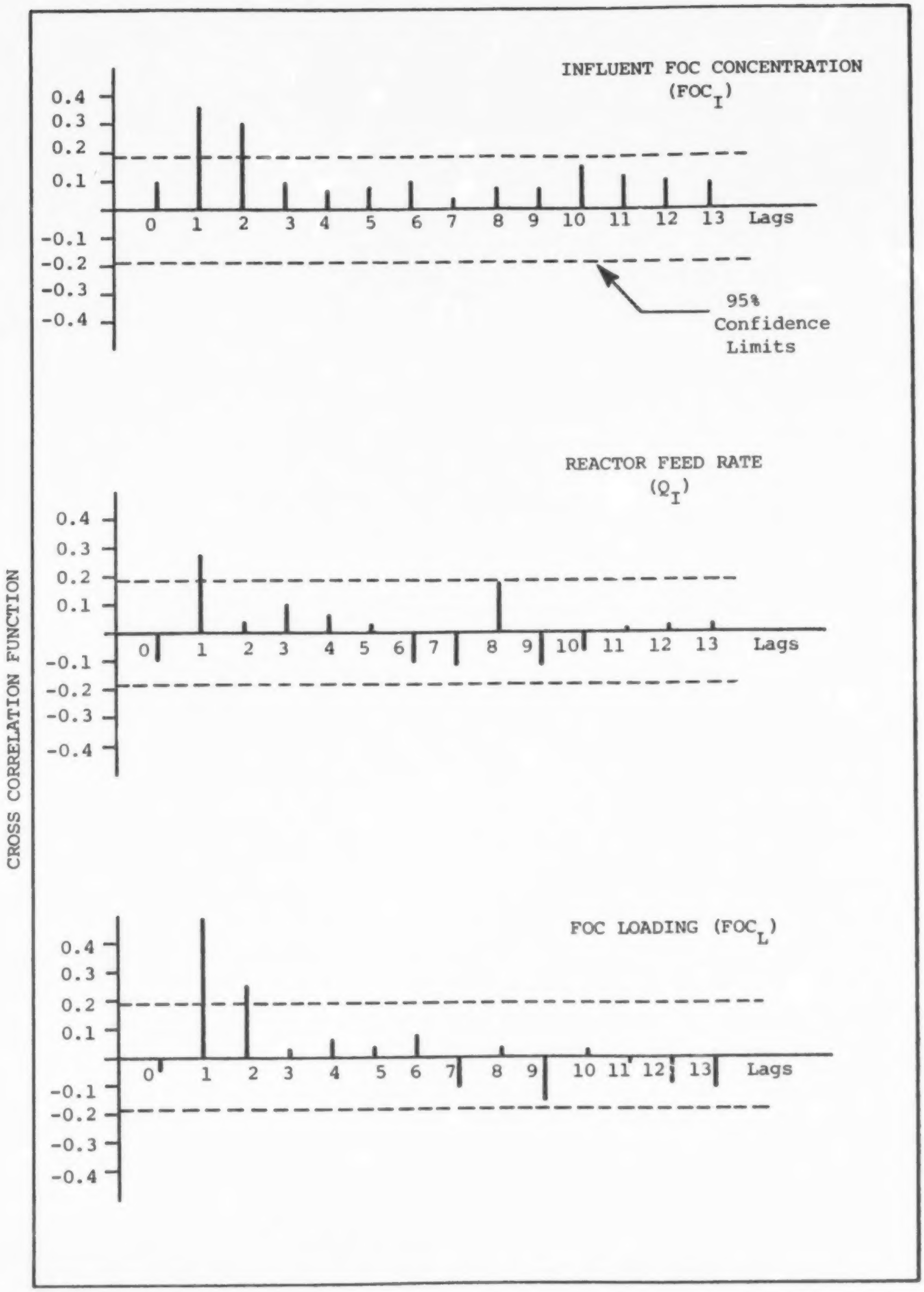


FIGURE 6.5 CROSS CORRELATION BETWEEN INFLUENT VARIABLES AND EFFLUENT FOC CONCENTRATION

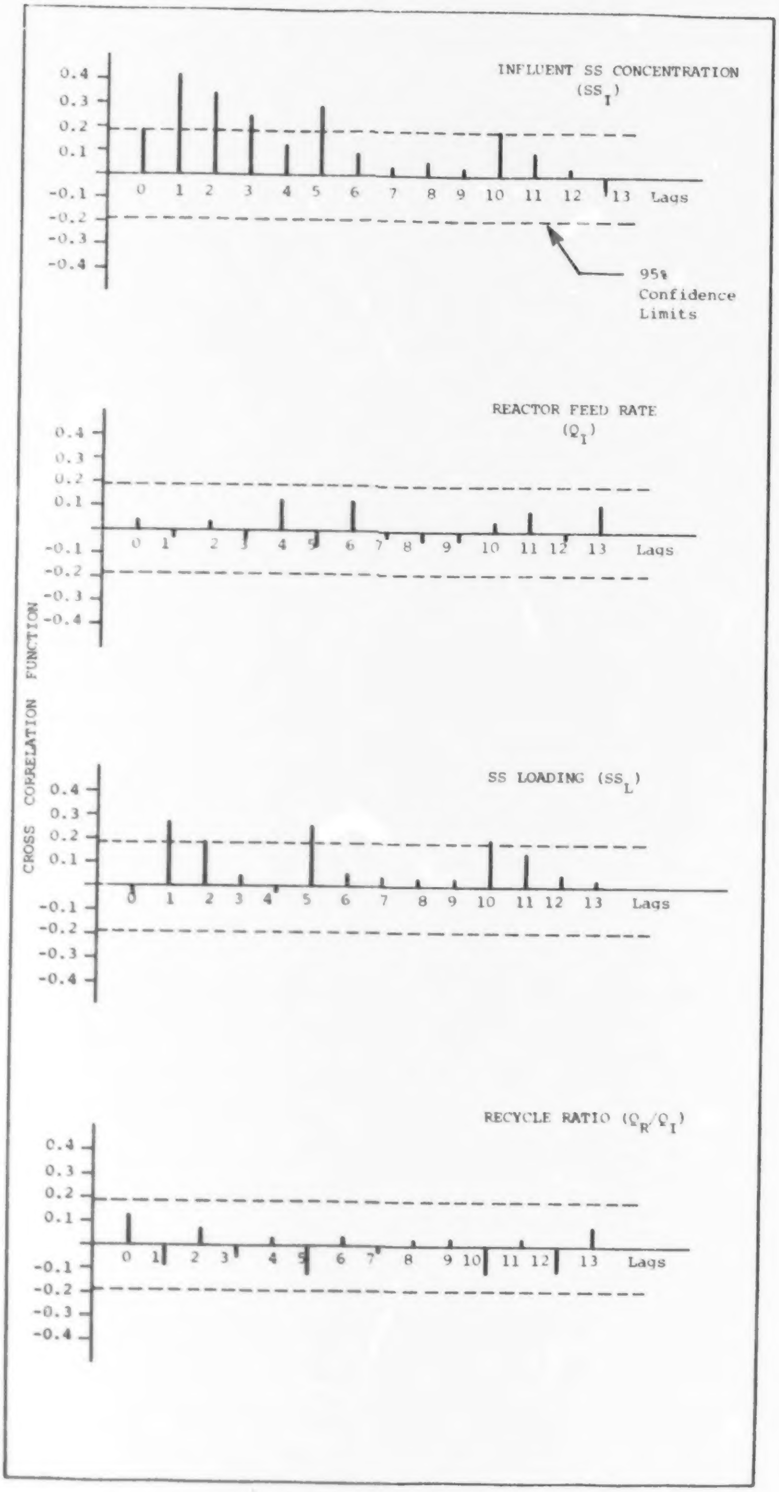


FIGURE 6.6 CROSS CORRELATION BETWEEN INFLUENT VARIABLES AND EFFLUENT SS CONCENTRATION

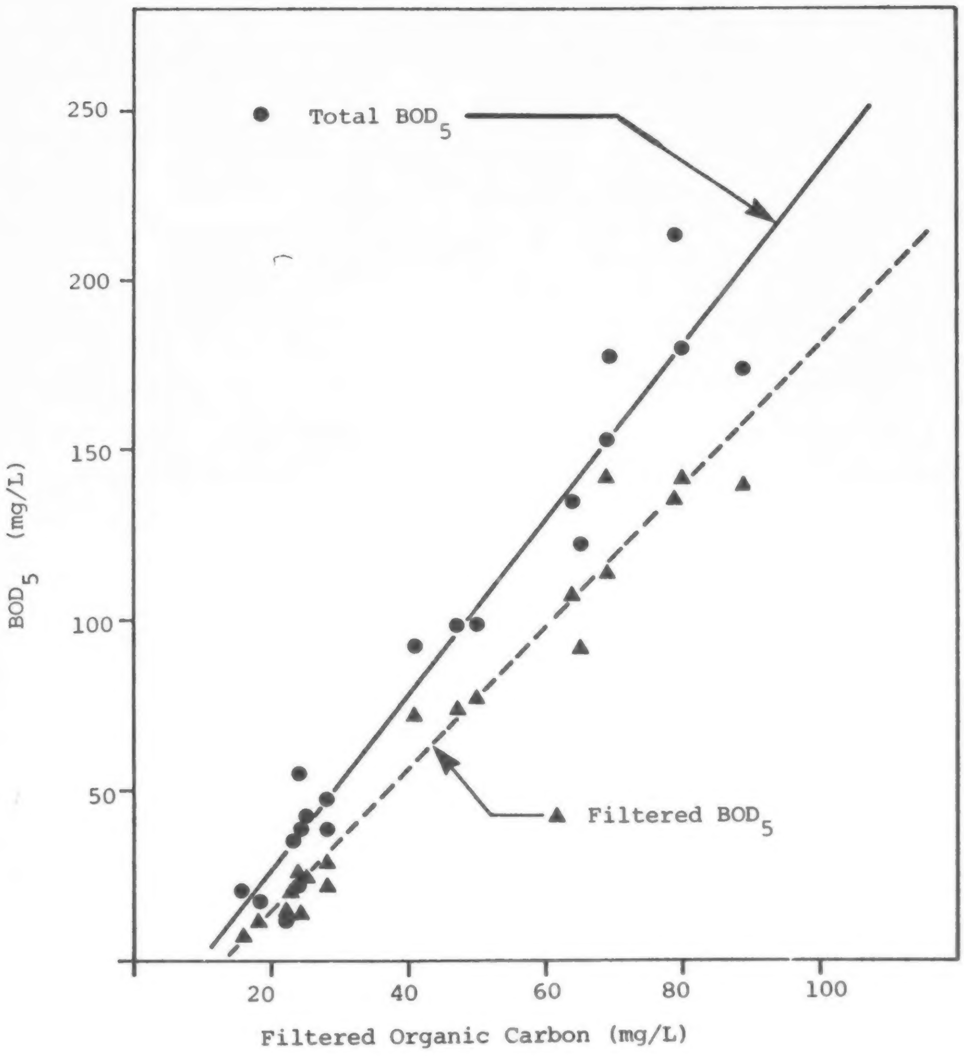


FIGURE 6.7 RELATIONSHIP BETWEEN BOD_5 AND ORGANIC CARBON CONCENTRATION

TABLE 6.1 SUMMARY OF CROSS-CORRELATION RESULTS

Input Variable	Output Response	Correlation Significance (95% Confidence)
FOC Concentration (FOC_I)	FOC Concentration (FOC_E)	Yes
FOC Load (FOC_L)	FOC Concentration (FOC_E)	Yes
Flow (Q_I)	FOC Concentration (FOC_E)	Yes
SS Concentration (SS_I)	SS Concentration (SS_E)	Yes
SS Load (SS_L)	SS Concentration (SS_E)	Yes
Flow (Q_I)	SS Concentration (SS_E)	No
Recycle Ratio (Q_R/Q_I)	SS Concentration (SS_E)	No

6.4 Development of Dynamic Models

Based on the measured responses, transfer function-noise models were developed relating the effluent response, in terms of FOC and suspended solids concentration, to the input variables found to be significantly correlated. Eight models, both single and multiple input, were developed and compared by an examination of the model residual sums of squares. Diagnostic checks on the model residuals verified that each model adequately described the experimental data.

The transfer function-noise models describing the FOC response are shown in Table 6.2.

The effuent filterable organic carbon response was best described by a three-variable multiple input model which combined the effects of influent filterable organic carbon concentration, FOC loading and flow. However, at the 95 percent confidence level, there was no significant difference between this and the FOC loading model. Based on an F-test at the same level of confidence, the FOC concentration and flow models were statistically inferior to the multiple input and loading models.

In the three-variable multiple input model for filterable organic carbon, the confidence limit on the parameter describing the effect of flow on the process response was centered around zero. Elimination of the flow term in the multiple input model and re-estimation of the parameters increased the magnitude of the residual sum of squares; the difference was not significant at the 95 percent confidence level. The resulting two-variable multiple input model, excluding the effect of flow, had about the same residual sum of squares as the FOC loading model.

TABLE 6.2 TRANSFER FUNCTION-NOISE MODELS FOR FOC RESPONSE

Input Variable	Model Form	Residual Sum of Squares
FOC Concentration (FOC_I)	$(FOC_E)_t = \frac{0.074B}{1-0.476B} (FOC_I)_t + \frac{(1-0.447B)}{(1-B)} a_t$	316
Flow (Q_I)	$(FOC_E)_t = 0.324B\,(Q_I)_t + \frac{(1-0.544B)}{(1-B)} a_t$	387
FOC Loading (FOC_L)	$(FOC_E)_t = \frac{0.0062B}{1-0.436B} (FOC_L)_t + \frac{(1-0.599B)}{(1-B)} a_t$	250
Multiple Input (FOC_I,Q_I,FOC_L)	$(FOC_E)_t = \frac{0.028B}{1-0.960B} (FOC_I)_t + \frac{0.0059B}{1-0.468B} (FOC_L)_t - 0.0466\,(Q_I)_t + \frac{1}{1-0.377B-0.399B^2} a_t$	238
Multiple Input (FOC_I,FOC_L)	$(FOC_E)_t = \frac{0.021B}{1-0.454B} (FOC_I)_t + \frac{0.0055B}{1-0.512B} (FOC_L)_t + \frac{1}{1-0.434B-0.457B^2} a_t$	249

The transfer function-noise models developed to describe the effluent suspended solids response are shown in Table 6.3. The response of the effluent suspended solids concentration was best described by a two-variable multiple input model combining the effect of the influent suspended solids concentration and the influent suspended solids mass loading. At the 95 percent confidence level, this was statistically better than either single input model. There was no statistically significant difference between the single input dynamic models for suspended solids.

6.5 Model Verification

All of the dynamic models developed were evaluated by comparing the predicted response to the response measured under the diurnal loading conditions (Section 5). The diurnal influent conditions of flow, concentration and loading were inserted in the transfer function models with noise terms (a_t) set to zero and the predicted fluidized bed effluent quality (FOC_E, SS_E) compared with the measured effluent quality. For the model verification, the influent and effluent data from June 20 (Figure 5.3) were used, as these data were collected under constant total hydraulic loading conditions applied during the development of the dynamic models.

The results of the model verification process for effluent filterable organic carbon are shown in Figure 6.8.

At the expense of including an additional variable (Q_I) and of estimating an additional parameter, the three-variable multiple input FOC transfer function model provided a better prediction of the effluent FOC concentration than the two-variable multiple input model. Although the three-variable model provided a better prediction, the parameter estimates for the two-variable model were more consistent with those for the single input models and appear to best describe the process dynamics. The single input transfer function models based on influent concentration, loading or flow, provided significantly poorer predictions. From this analysis, filterable organic carbon removal is best described by a relationship which accounts for the effect of FOC concentration and mass loading.

The results of the model verification process for effluent suspended solids are shown in Figure 6.9. During the time series experimentation, there was no return of sand from the separator to the reactor. Therefore, the peak effluent solids concentrations during the diurnal loading studies were not included in the dynamic modelling. For model verification, the effect of the sand return operation on the effluent suspended solids

TABLE 6.3 TRANSFER FUNCTION-NOISE MODELS FOR SS RESPONSE

Input Variable	Model Form	Residual Sum of Squares
SS Concentration (SS_I)	$(SS_E)_t = \frac{0.082B}{1-0.390B}(SS_I)_t + a_t$	1604
SS Loading (SS_L)	$(SS_E)_t = \frac{0.0046B}{1-0.430B}(SS_L)_t + a_t$	1692
Multiple Input (SS_I, SS_L)	$(SS_E)_t = \frac{0.067B}{1-0.559B}(SS_I)_t + \frac{0.0014B}{1-0.884B}(SS_L)_t + \frac{1}{1-0.087B}a_t$	1406

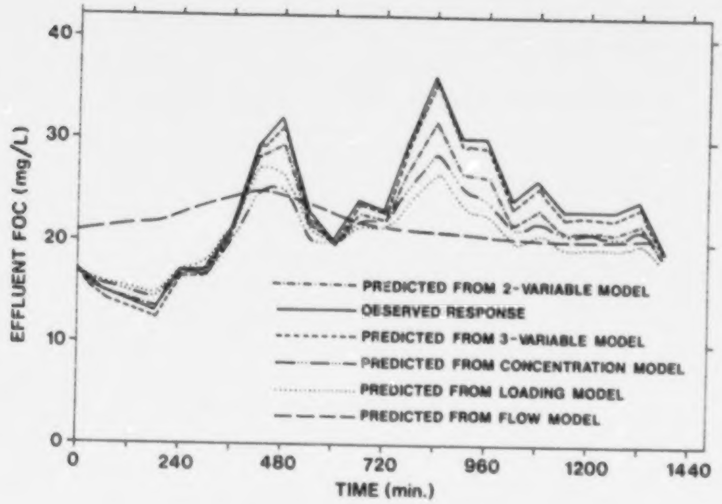


FIGURE 6.8 COMPARISON OF PREDICTED AND OBSERVED RESPONSE FOR EFFLUENT FOC

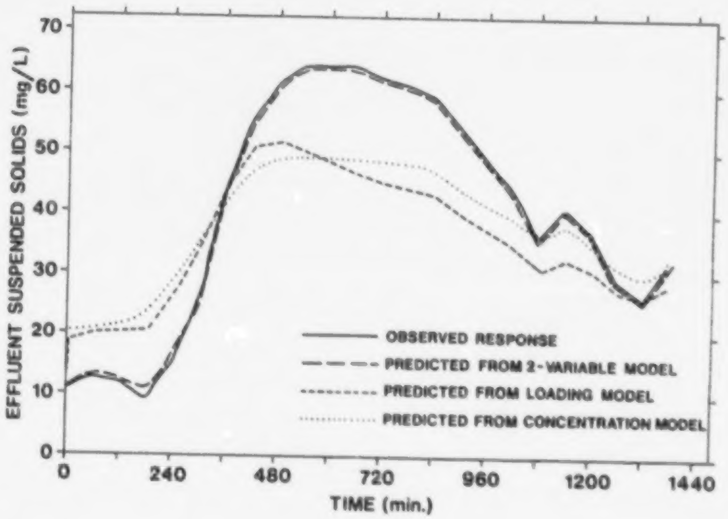


FIGURE 6.9 COMPARISON OF PREDICTED AND OBSERVED RESPONSE FOR EFFLUENT SUSPENDED SOLIDS

concentration was eliminated from the June 20 experimental data by interpolating a smooth curve to represent the concentration between approximately 1500 and 1700 hours.

The individual dynamic models, based solely on influent suspended solids concentration or influent solids loading, showed similar deficiencies in predicting the response of the effluent suspended solids concentration, as was evident in the prediction of the effluent FOC concentration by single input models. Both single input transfer function models underestimated the process gain under peak loading. The multiple input dynamic model significantly improved the accuracy of the predicted process response. It is concluded that effluent suspended solids concentration also is dependent upon the concentration and mass loading of solids.

6.6 Summary

The time series analysis of the process indicated that the effluent quality in terms of filterable organic carbon concentration is most significantly affected by the influent FOC concentration and the process FOC loading. Hydraulic loading (Q_I) has a lesser impact. The effluent suspended solids concentration is also controlled by the influent concentration and the process SS loading.

These results, in general, agree with the findings of the pseudo-steady state investigations (Section 4). Of the variables investigated under pseudo-steady state conditions, the effluent total BOD_5 and filtered BOD_5 were most significantly correlated to the respective influent concentrations. The process loading and hydraulic retention time (HRT) were also significantly correlated to the effluent quality; of these, loading had a more pronounced effect. Based on the dynamic modelling, the effluent suspended solids concentration was most significantly affected by the influent concentration. Under steady state conditions, the effluent quality was also affected by the process loading and the hydraulic retention time.

A direct comparison with the reported responses of other biological systems modelled at the WTC using similar procedures(38,39) is difficult. No modelling of the responses of the effluent suspended solids from the activated sludge or rotating biological contactor (RBC) processes was conducted.

The response of the effluent filterable organic carbon concentration in both cases was modelled solely on the process FOC loading, since the reported dynamic models based on the influent FOC concentration were found to be less precise in predicting the effluent response. No multiple input models were reported. Model verification for the

RBC was based on the combined transfer function-noise model, whereas only the transfer function portion of the model identifying the process dynamics was used for the fluidized bed. Use of the transfer function gives a better indication of the ability of the model to emulate the underlying process dynamics.

The steady state process gains reported(39) for the activated sludge system and the RBC loading models for organic carbon removal from domestic wastewaters were 0.026 and 0.048 min/L, respectively with the RBC responding to a greater extent, and more rapidly. As a basis for comparison, the gain for the single input FOC loading model for the fluidized bed was approximately 0.011 min/L, indicating a lesser response to a unit change in process loading than occurs in either the activated sludge or RBC systems. The single input gain was the same as the gain of 0.011 min/L associated with the loading term of the two-variable multiple input model. However, the fluidized bed single input FOC loading model underestimated the process gain due to the additional impact of influent FOC concentration on the effluent response. The process gain based on the single input influent concentration model was approximately 0.14 compared to 0.20 for the RBC. For the multiple input case with the model parameters re-estimated, the gain due to the concentration term was reduced to 0.038 but the estimated parameter was relatively imprecise. In operational situations, the increased process loading results from increases in both influent concentration and influent flow rate, and both factors must be taken into account in predicting the response.

It is also informative to estimate the time constants associated with the two-variable model where, for the first-order systems identified here with eight-minute sampling interval, the time constant (τ) in minutes is defined by:

$$\tau \quad = \quad -8.0/\ln \delta_1$$

For the FOC concentration and loading terms, the time constant is 10.1 and 12.0 minutes, respectively. This indicates that 63.2 percent of the fluidized bed response to a step change in input concentration will occur within 10.1 minutes and that 95 percent response (i.e., essentially complete response) will occur in about three time constants or 30.3 minutes with an exponential approach to the ultimate steady state value. A similar analysis applies to the loading term. Since the discrete transfer function models identified for the RBC and activated sludge process(39) were of zero-order, their time constants cannot be identified, but it is anticipated that these would be significantly longer than for the fluidized bed.

7 FLUIDIZED BED EFFLUENT SOLIDS CHARACTERISTICS AND TREATABILITY

As discussed in detail in Section 4, and as observed by other researchers involved in the assessment of fluidized bed technology(10,25), the design loading on the process is controlled by the fluidized bed's ability to remove suspended solids. Residual suspended solids in the treated effluent contribute a significant fraction of the final effluent total BOD_5. Therefore, information regarding the nature and the treatability of the effluent suspended solids is important to the application of fluidized bed technology.

7.1 Fluidized Bed Effluent Solids Characteristics

In such suspended growth systems as activated sludge, the effluent suspended solids are considered to be biomass lost from the biological reactor through the secondary clarifier. However, Murphy and Wilson(34) demonstrated that, based on the oxygen uptake associated with the influent, effluent and biomass solids from an RBC, the effluent solids were composed primarily of influent solids not removed in the reactor. As the fluidized bed system is also a fixed film biological treatment process, residual influent solids may represent a large fraction of the residual solids in the treated effluent.

To evaluate the nature of the fluidized bed effluent suspended solids, the adenosine triphosphate (ATP) content of feed solids, effluent solids and reactor biomass were compared. The ATP content of the solids provides a qualitative indication of the fraction of viable biomass present(35). The analytical procedure has been described elsewhere(36).

The ATP contents of the influent, effluent and fluidized bed biomass are shown as a probability plot in Figure 7.1 and the results summarized in Table 7.1. At the 95 percent confidence level, all three sources of particulate matter contained significantly different ATP contents per milligram suspended solids. The ATP content of the effluent solids is significantly lower than that of the biomass. These results would indicate a significant quantity of influent particulate matter passes through the fluidized bed reactor and is measured as effluent suspended solids. A similar conclusion has been drawn by Sutton _et al_(9) based on the oxygen uptake associated with the influent and effluent solids and the reactor biomass.

The ATP and suspended solids data for the effluent from the fluidized bed indicate a trend toward a higher ATP content per unit mass of suspended solids as the

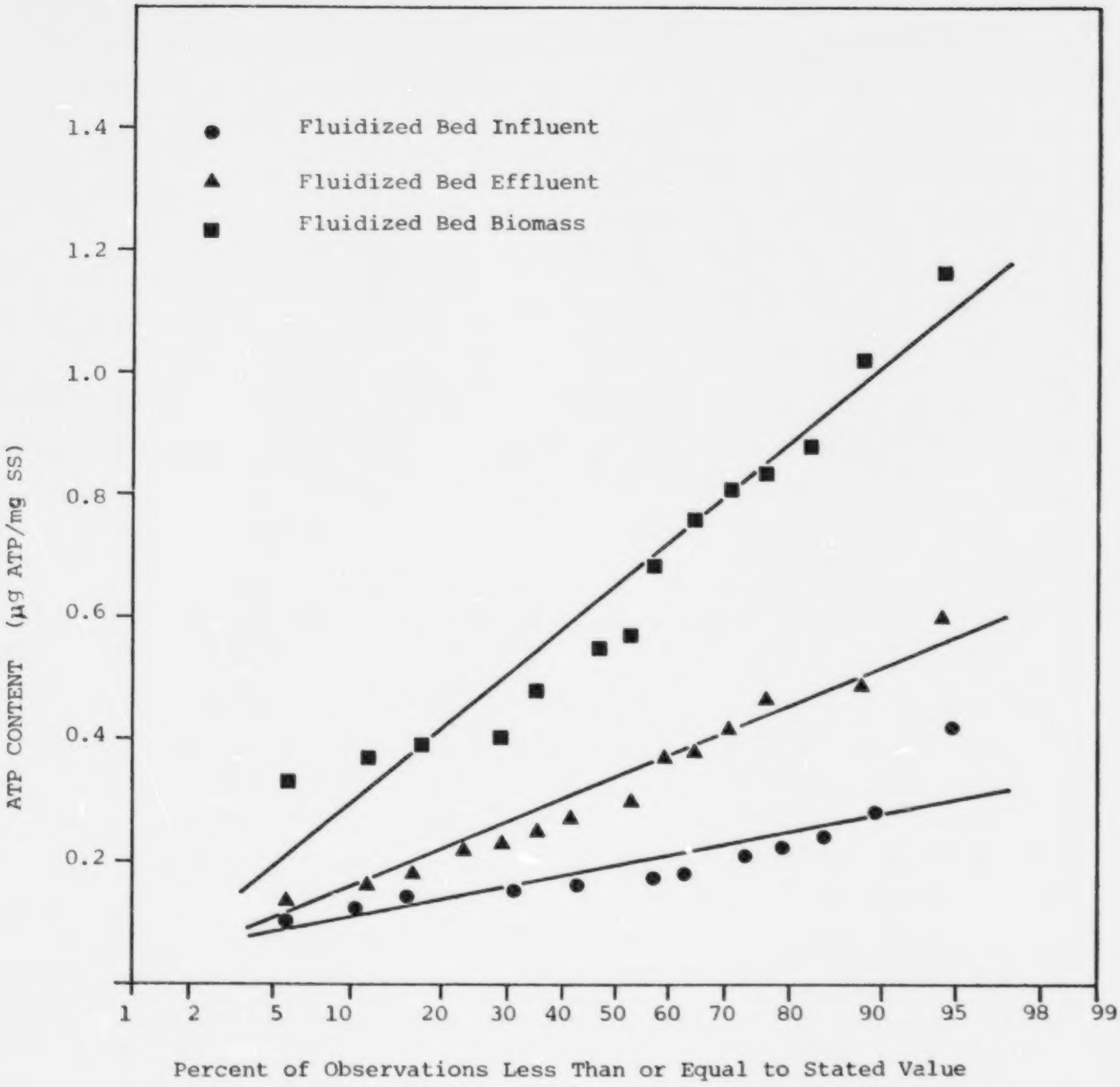


FIGURE 7.1 COMPARISON OF INFLUENT, EFFLUENT AND BIOMASS ATP CONTENT

TABLE 7.1 COMPARISON OF THE ATP CONTENT OF INFLUENT, EFFLUENT AND REACTOR SOLIDS

Source of Particulate Matter	Number of Observations	ATP Content (μg/mg SS)	
		Median	90%*
Fluidized Bed Influent	18	0.19	0.28
Fluidized Bed Effluent	16	0.34	0.52
Fluidized Bed Biomass	16	0.65	1.01

* 90 percent of observations less than or equal to stated value.

concentration of effluent suspended solids decreases (Figure 7.2). Despite the considerable scatter in these data, it would appear that biological solids from the fluid bed represent a larger fraction of the total effluent solids as the concentration of effluent solids decrease.

During the fluidized bed sampling and analyses, an attempt was made to estimate the effect of fluidized bed operating conditions on the characteristics, in terms of ATP, of the effluent solids. However, there are insufficient data to draw any conclusions.

Although considerable research into the effects of various factors on the ATP content of bacterial cells would be necessary to confirm the validity of directly comparing the ATP content of the influent and effluent streams, a simple mass balance on the ATP indicates that the influent solids represent an average of approximately 65 percent of the effluent solids over the range of operating conditions of the fluidize bed.

7.2 Fluidized Bed Effluent Solids Treatability

7.2.1 Sedimentation. Effluent samples from the fluidized bed system were routinely settled, under quiescent conditions for 30 minutes, and the supernatant quality monitored. The effect on the effluent suspended solids concentration is shown in Figure 7.3, and the effect on the effluent BOD_5 in Figure 7.4. The quality of the unsettled and settled fluidized bed effluent is compared in Table 7.2. These data represent the quality of the fluidized bed effluent over the range of operating conditions included in the steady state experimental design (Section 4).

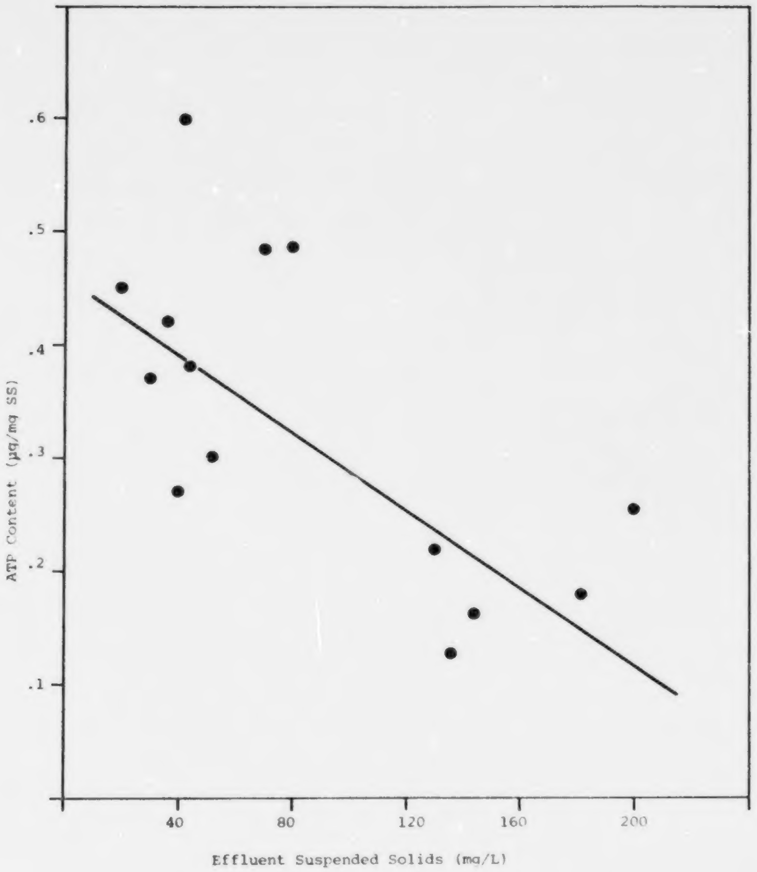


FIGURE 7.2 RELATIONSHIP BETWEEN EFFLUENT SUSPENDED SOLIDS CONCENTRATION AND ATP CONTENT

Thirty minutes of quiescent settling was insufficient to improve the fluidized bed effluent quality to meet a standard of 15 mg/L BOD_5 and 15 mg/L suspended solids.

7.2.2 Chemical Coagulation. A brief evaluation of the potential for chemical coagulation of the fluid bed effluent solids was conducted. Jar tests were carried out once at the Dorr-Oliver Oxitron pilot plant in Orillia. This pilot-scale system has been described in detail elsewhere(9,25,26).

During the jar testing, a number of chemical coagulants were assessed, including:

1) high molecular weight acrylamide cationic copolymer (Aquafloc 6461),
2) low molecular weight, high charge polyamine cationic polymer (Aquafloc 431),
3) low molecular weight, high charge cationic polyamine in conjunction with an anionic polyacrylamide (Aquafloc 6466),
4) alum along with a variety of anionic polymers,

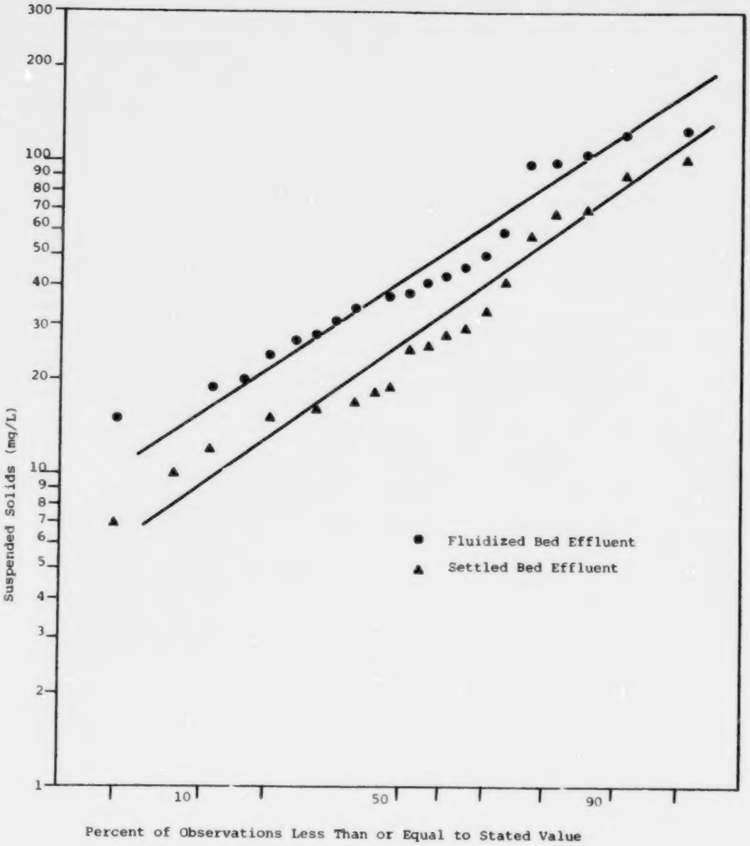


FIGURE 7.3 EFFECT OF SETTLING ON FLUIDIZED BED EFFLUENT SUSPENDED SOLIDS

5) pickle liquor (plus sodium hypochloride to convert ferrous iron to ferric iron and alkalinity to neutralize free acid) in conjunction with a moderately charged anionic polymer, and

6) ferric chloride in combination with anionic polyacrylamides with a range of charge densities.

The results were evaluated qualitatively.

The optimum treatment approach involved adding 10 mg/L ferric chloride (as $FeCl_3$) and anionic polyacrylamide (Aquafloc 6466, 20 percent charge density) at a dosage of 0.1 to 0.2 mg/L. Further testing would be necessary to precisely define the solids settling rate and treated effluent quality.

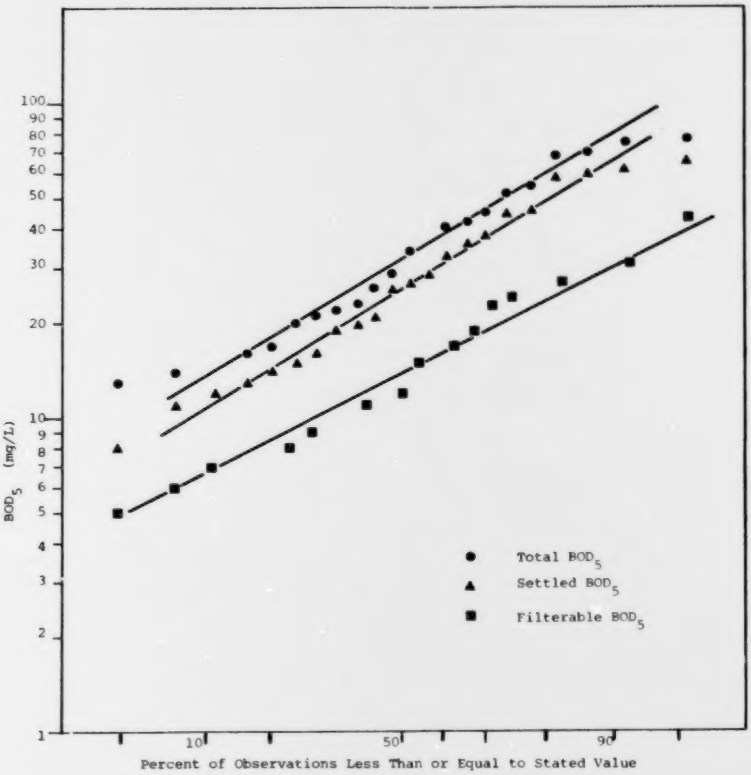


FIGURE 7.4 EFFECT OF SETTLING ON FLUIDIZED BED EFFLUENT BOD_5

TABLE 7.2 EFFECT OF SETTLING ON FLUIDIZED BED EFFLUENT QUALITY

	Unsettled Fluidized Bed Effluent		Settled Fluidized Bed Effluent	
Parameter (mg/L)	Median	90%*	Median	90%*
Total BOD_5	32	81	26	66
Filtered BOD_5	14	30	-	-
Suspended Solids	42	116	26	78

* 90 percent of observations less than or equal to stated value.

7.2.3 Granular Media Filtration. An intensive experimental program was conducted to compare the filtration characteristics of treated effluents from the pilot system and from a conventional extended aeration activated sludge biological treatment system using granular media gravity filters. The parameters evaluated were:

1) type of media
2) depth of media
3) hydraulic loading
4) influent solids concentration

Description of Biological Treatment Systems: Operating conditions of the reactor were varied over a range of recycle ratios and hydraulic retention times so that the effluent suspended solids concentration would be varied over an appropriate range for the design of filtration experimental runs. The effluent dissolved oxygen concentration was maintained at 5 mg/L.

During the experimental runs to define optimal media type and depth, the fluidized bed operating conditions were as follows:

Hydraulic Retention Time:	20 minutes
Recycle Ratio:	2.5
Total Hydraulic Flux:	0.51 $m^3/m^2 \cdot min$.

The operational conditions for the fluidized bed process are summarized in Table 7.3.

TABLE 7.3 SUMMARY OF FLUIDIZED BED OPERATING CONDITIONS

	Fluid Bed Operating Conditions		
Desired Range of SS Concentration (mg/L)	HRT* (minutes)	Recycle Ratio (Q_R/Q_I)	Hydraulic Flux** ($m^3/m^2 \cdot min$)
>30	10	2.0	0.87
≃30	20	2.5	0.51
<30	30	3.0	0.39

* hydraulic retention time based on primary clarifier effluent flowrate (Q_I)

** hydraulic flux based on total flow ($Q_I + Q_R$)

The single modification to the procedure during the evaluation of effluent filterability compared to the procedure for the remainder of the steady state evaluation of fluidized bed performance related to the return of sand from the biomass wasting and sand cleaning system. The return of sand once per day to the fluidized bed resulted in a high effluent suspended solids concentration for a short time. This would not be representative of a process designed to return sand and waste biomass more frequently; so during filter operation, no sand was returned except between filtration runs.

The biological reactor of the extended aeration plant was a baffled 131.7 m^3 vessel. Aeration and mixing were provided by a diffused air system. The reactor was operated at a flow of 90.8 to 113.5 L/min, equivalent to a nominal hydraulic retention time ranging from 19.3 to 24.1 hours. The biological system was operated at a mean cell residence time (SRT) of approximately five days and mixed liquor suspended solids concentration of approximately 1000 mg/L.

The hydraulic loading on the 10.9 m^2 secondary clarifier was in the range from 12.0 to 15.0 $m^3/m^2 \cdot d$. The relatively low hydraulic loading of the extended aeration plant resulted in a high quality effluent (mean effluent suspended solids concentration of approximately 6 mg/L). To evaluate the filterability of the extended aeration plant effluent over a range comparable to the fluidized bed plant effluent, mixed liquor from the aeration tank was added to the clarifier overflow at a controlled rate by a variable-speed pump.

The extended aeration system was fed degritted raw sewage directly from the Burlington Skyway Water Pollution Control Plant. The fluidized bed system was operated on stored sewage so that the feed quality and effluent quality would be relatively constant. The feed underwent primary clarification prior to entering the reactor.

Both biological processes were operated at ambient temperature. The mean temperature of the fluidized bed effluent was 21.1°C (range 20°C to 22°C) compared to the extended aeration effluent mean temperature of 21.0°C (range 20°C to 22°C).

Description of Granular Media Gravity Filters: Two similar downflow gravity filters were constructed and installed to compare the filterability of extended aeration plant effluent and fluidized bed effluent. The filters were fed by gravity from the respective biological treatment systems.

A schematic diagram of the filter system is shown in Figure 7.5. The filters, 10.2 cm in diameter and approximately 3.0 m high, were operated in the constant

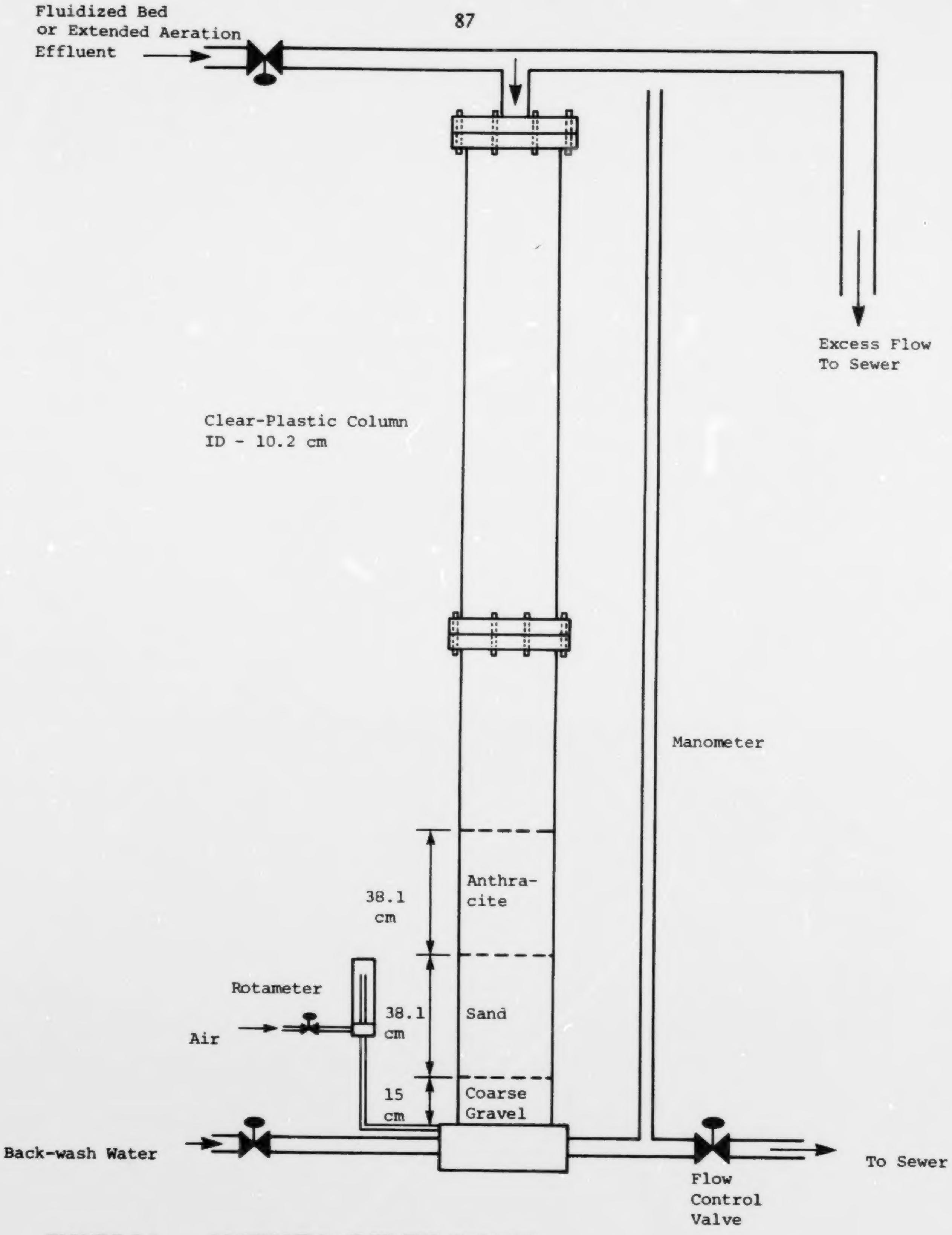

Fluidized Bed
or Extended Aeration
Effluent
Excess Flow
To Sewer
Clear-Plastic Column
ID - 10.2 cm
Manometer
38.1
cm
Anthra-
cite
Rotameter
Air
38.1
cm
Sand
15
cm
Coarse
Gravel
Back-wash Water
To Sewer
Flow
Control
Valve

FIGURE 7.5 SCHEMATIC OF FILTER COLUMN

pressure-declining rate mode. The flow rate through the filter was controlled by a manual diaphragm valve on the filtrate. A perforated plate and 15 cm of pea gravel were used as support for the filtration media.

After the addition of the appropriate quantity of media to the filter, the flow rate was set by the effluent flow control valve. During the filtration run, the flow rate and headloss through the filter were measured hourly or half-hourly depending on the anticipated length of the run. Grab samples of filter influent and filtrate were collected for analysis of suspended solids concentration and turbidity. Composite samples of the filter influents and effluents were also collected for BOD_5 and suspended solids analyses.

At the completion of the filter run, the filter media were backwashed to expand the media approximately 70 percent, and air scoured at approximately 0.014 m^3/min for 10 to 15 minutes. The objective of the backwash was to ensure that the filter media were completely cleaned and there was no progressive deterioration in filtration characteristics from accumulation of solids in the bed. Therefore, no quantitative measurement of backwash requirements were made. Visual observations of the nature of the backwash solids and the location of solids in the filter bed were recorded.

Experimental Design: The experimental program was divided into two phases. During Phase I, the optimal media type and depth for filtration of fluidized bed effluents were selected at constant hydraulic and solids loading conditions. In Phase II, the filter performance was evaluated over a range of hydraulic and solids loading conditions using the optimum filter configurations selected from Phase I.

The Phase I filtration runs, for both the fluidized bed effluent and the extended aeration plant effluent, were conducted at an initial hydraulic loading of 162.8 L/min•m^2.

The desired effluent suspended solids concentration for the fluidized bed effluent was from 30 to 50 mg/L. For the extended aeration effluent filter, clarifier overflow was used directly without added mixed liquor solids; the filter influent had a low suspended solids concentration, in the range from 5 to 7 mg/L.

Five filter configurations were evaluated during Phase I:

Filter Configuration	Media Type	Media Depth (cm)
1	Sandblast Sand	38.1
2	Sandblast Sand	76.2
3	Fluid Bed Sand	38.1
4	Fluid Bed Sand	76.2
5	Dual Media Anthracite/Sand	76.2*

*38.1 cm of Anthracite and 38.1 cm of Sand.

The particle size distributions are shown in Figure 7.6 (Sandblast Sand), Figure 7.7 (Fluid Bed Sand) and Figure 7.8 (Anthracite). The fluid bed sand was the same medium used in the biological reactor of the process. Parameters characterizing the filtration media are summarized in Table 7.4.

Based on the Phase I investigations, the selected filter configuration was evaluated over a range of hydraulic loadings and influent suspended solids concentrations. Six filtration runs on each type of effluent (fluidized bed and extended aeration plant) were conducted.

Hydraulic loading conditions evaluated were 81.4 L/min•m^2, 162.8 L/min•m^2 and 244.2 L/min•m^2. For the fluidized bed effluent, a range of suspended solids concentrations from about 20 mg/L to 60 mg/L was assessed. For the extended aeration plant effluent, the suspended solids concentration, augmented with mixed liquor solids, ranged from 24 mg/L to 43 mg/L.

The experimental design for Phase II is summarized in Table 7.5. Despite efforts to control the suspended solids concentration of the effluents within fixed ranges, the difference between desired and actual effluent qualities shows that this was not completely successful.

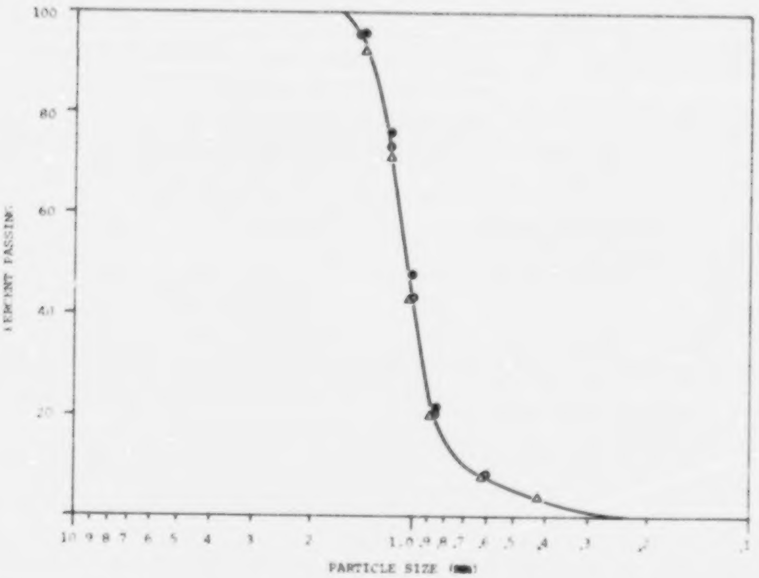


FIGURE 7.6 PARTICLE SIZE CHARACTERIZATION OF SANDBLAST SAND

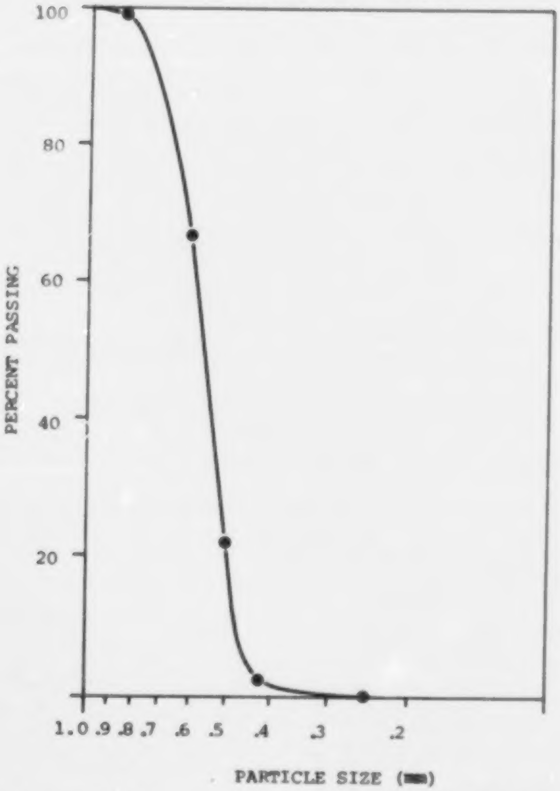


FIGURE 7.7 PARTICLE SIZE CHARACTERIZATION OF FLUIDIZED BED SAND

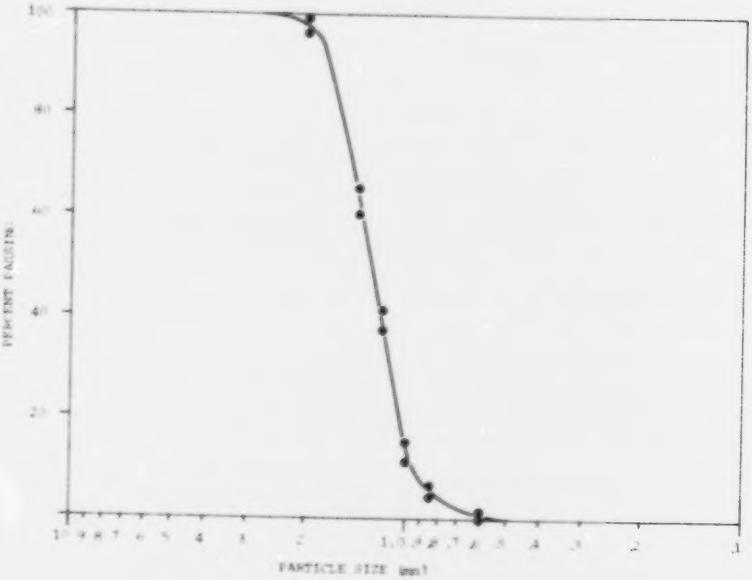


FIGURE 7.8 PARTICLE SIZE CHARACTERIZATION OF ANTHRACITE

TABLE 7.4 CHARACTERISTICS OF FILTRATION MEDIA

Medium	Effective Size (d_{10}, mm)	Uniformity Coefficient (d_{60}/d_{10})
Sandblast Sand	0.70	1.57
Fluid Bed Sand	0.48	1.23
Anthracite	0.96	1.41

TABLE 7.5 EXPERIMENTAL DESIGN OF FILTRATION RUNS FOR PHASE II

Run No.	Initial Hydraulic Loading (L/min•m^2)	Suspended Solids Concentration (mg/L) Fluid Bed Effluent Desired	Fluid Bed Effluent Actual*	Extended Aeration Effluent Desired	Extended Aeration Effluent Actual*
1	81.4	>30	58	>30	27
2	244.2	>30	54	>30	30
3	81.4	<30	41	<30	43
4	244.2	<30	22	<30	24
5	162.8	≃30	62	≃30	32
6	162.8	≃30	32	≃30	34

*Based on composite sample of influent to filters for duration of filter run.

Phase I Results - Selection of Optimal Media Type and Depth: The effect on filtration rate and filter headloss on filter run time for the fluidized bed effluent with the various filter configurations are shown in Figures 7.9 and 7.10, respectively. Similar data for the extended aeration plant effluent are in Figures 7.11 and 7.12.

From Figures 7.9 and 7.10, the longest filter runs were attained with the dual media configuration. Conversely, the fluidized bed sand (d_{10} = 0.48 mm) produced the shortest filter runs. In the case of the single medium sand filters, increasing the medium depth reduced the length of the filter runs with some improvement in filtrate quality. The performance of the filters is summarized in Table 7.6. In all cases, effluent suspended solids concentrations were reduced to less than 15 mg/L from the influent levels of 33 to 52 mg/L. There was no indication of deterioration in filtrate quality as the run progressed. Effluent total BOD_5 was also reduced to less than 15 mg/L. The filtrate BOD_5 concentration was primarily dependent on the soluble BOD_5 concentration in the effluent.

Similarly, during filtration of extended aeration plant effluents, the fluidized bed sand produced the shortest filtration runs (Figures 7.11 and 7.12). The coarser sand (sandblast sand, d_{10} = 0.70 mm) and the dual media filters provided comparable performance. Long filtration runs (greater than 24 hours) resulted, due to the extremely low

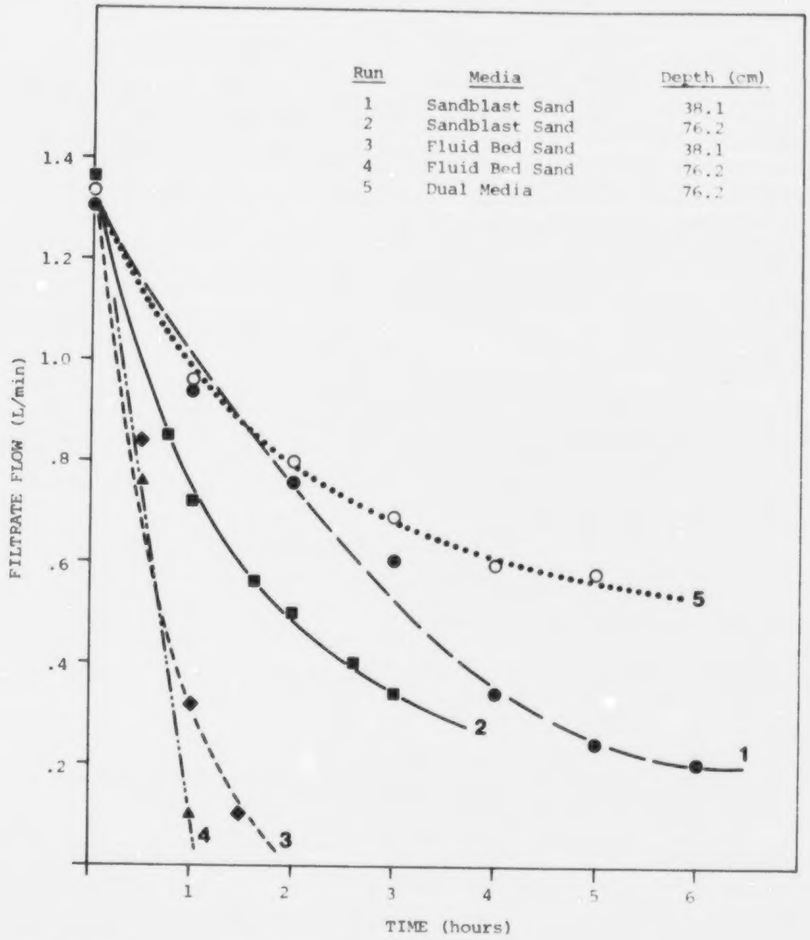


FIGURE 7.9 EFFECT OF FILTER CONFIGURATION ON FILTER RUN LENGTH - FLUIDIZED BED EFFLUENT

suspended solids loadings. The filter performance on extended aeration plant effluents is summarized in Table 7.7.

The Phase I investigations indicated that the dual media filter provided the best performance for the fluidized bed effluents. This filter configuration was also as efficient as the single medium filter containing the coarser sand (d_{10} = 0.70 mm) for the extended aeration plant effluents. However, during backwashing and air scouring of the dual media filters, there was considerable intermixing of the fine fluidized bed sand (d_{10} = 0.48 mm) and the anthracite. This problem was largely eliminated using coarser sand with anthracite. For the Phase II investigations, a dual media filter containing

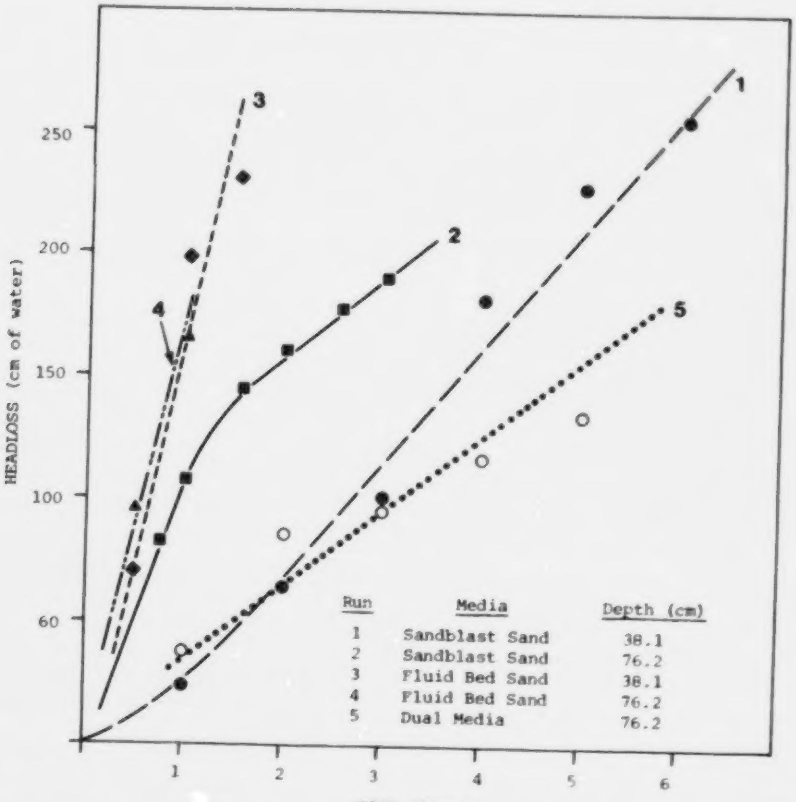


FIGURE 7.10 EFFECT OF FILTER CONFIGURATION ON FILTER HEADLOSS - FLUIDIZED BED EFFLUENT

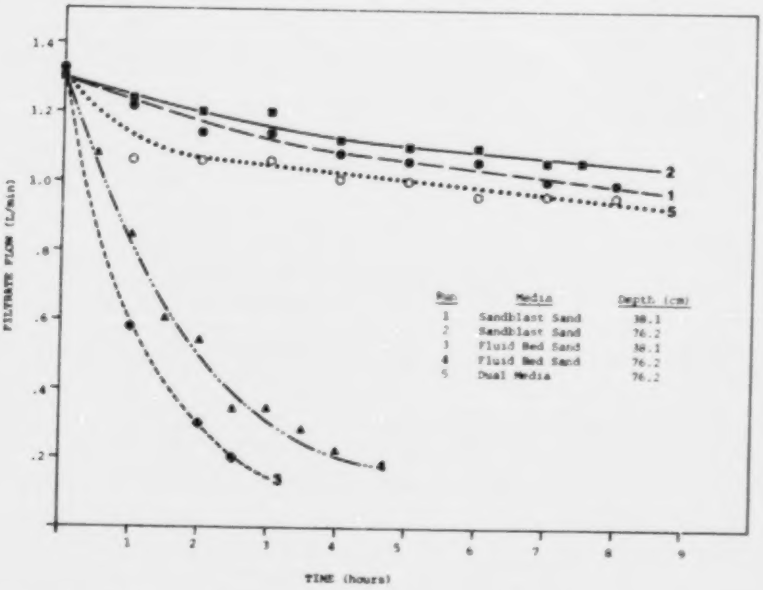


FIGURE 7.11 EFFECT OF FILTER CONFIGURATION ON FILTER RUN LENGTH - EXTENDED AERATION PLANT EFFLUENT

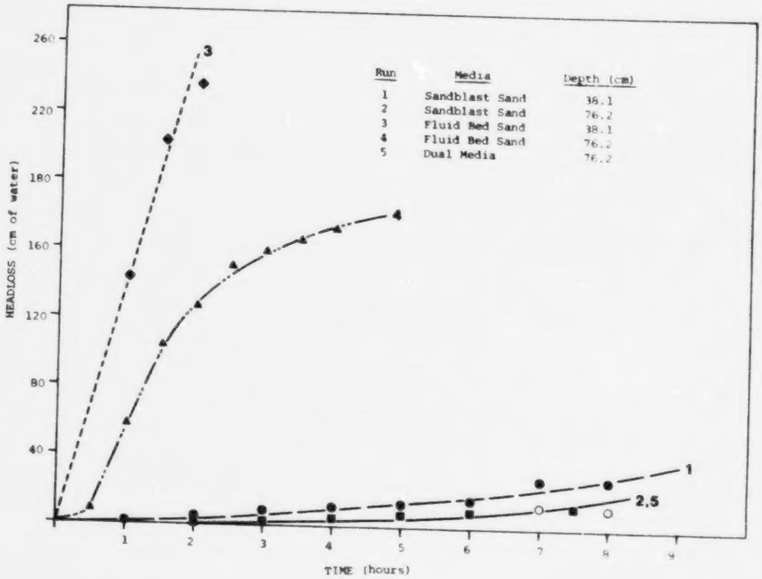


FIGURE 7.12 EFFECT OF FILTER CONFIGURATION ON FILTER HEADLOSS - EXTENDED AERATION PLANT EFFLUENT

38.1 cm of anthracite and 38.1 cm of sandblast sand (d_{10} = 0.70 cm) was selected as the better filter for both secondary effluents.

Phase II Results — Evaluation of Filter Performance: The results of the filtration of fluidized bed effluents and extended aeration plant effluents for Phase II are shown graphically in Figures 7.13 to 7.18, in terms of the effect of filter run time on filtrate flow rate and filter headloss. The results are summarized in Table 7.8.

Substitution of the coarser sand in the dual media configuration (d_{10} = 0.70 mm in Phase II as opposed to d_{10} = 0.48 mm in Phase I) significantly increased the fluidized bed filter run time. At comparable hydraulic loadings (162.8 L/min•m^2), filter runs longer than 24 hours were possible with the coarser sand and anthracite, compared to seven hours with the fine sand in Phase I. There was some deterioration in effluent quality.

TABLE 7.6 SUMMARY OF FILTER PERFORMANCE ON FLUIDIZED BED EFFLUENTS

Run No.	Media Type	Depth (cm)	Suspended Solids Concentration* (mg/L)		BOD$_5$ Concentration* (mg/L)				Estimated Run Length (hours) to 200 cm Headloss
					Influent		Effluent		
			Influent	Effluent	Total	Soluble	Total	Soluble	
1	Sandblast Sand	38.1	48	9	24	6	7	6	4.75
2	Sandblast Sand	76.2	52	6	31	10	10	9	3.25
3	Fluid Bed Sand	38.1	33	6	24	11	11	8	1.15
4	Fluid Bed Sand	76.2	45	3	26	8	11	6	1.15
5	Dual Media	76.2	44	2	30	11	11	10	6.5**

* Based on influent and effluent composite samples.

** Estimated based on extrapolation of experimental data.

TABLE 7.7 SUMMARY OF FILTER PERFORMANCE ON EXTENDED AERATION PLANT EFFLUENT

Run No.	Media Type	Depth (cm)	Suspended Solids Concentration* (mg/L)		BOD_5 Concentration* (mg/L)				Estimated Run Length (hours) to 200 cm Headloss
			Influent	Effluent	Influent		Effluent		
					Total	Soluble	Total	Soluble	
1	Sandblast Sand	38.1	7	2	6	2	3	3	>24**
2	Sandblast Sand	76.2	5	3	9	7	5	4	>24**
3	Fluid Bed Sand	38.1	5	2	7	7	6	4	1.5
4	Fluid Bed Sand	76.2	6	4	5	4	5	4	∽ 8**
5	Dual Media	76.2	6	4	7	4	4	3	>24**

* Based on influent and effluent composite samples.

** Estimated based on extrapolation of experimental data.

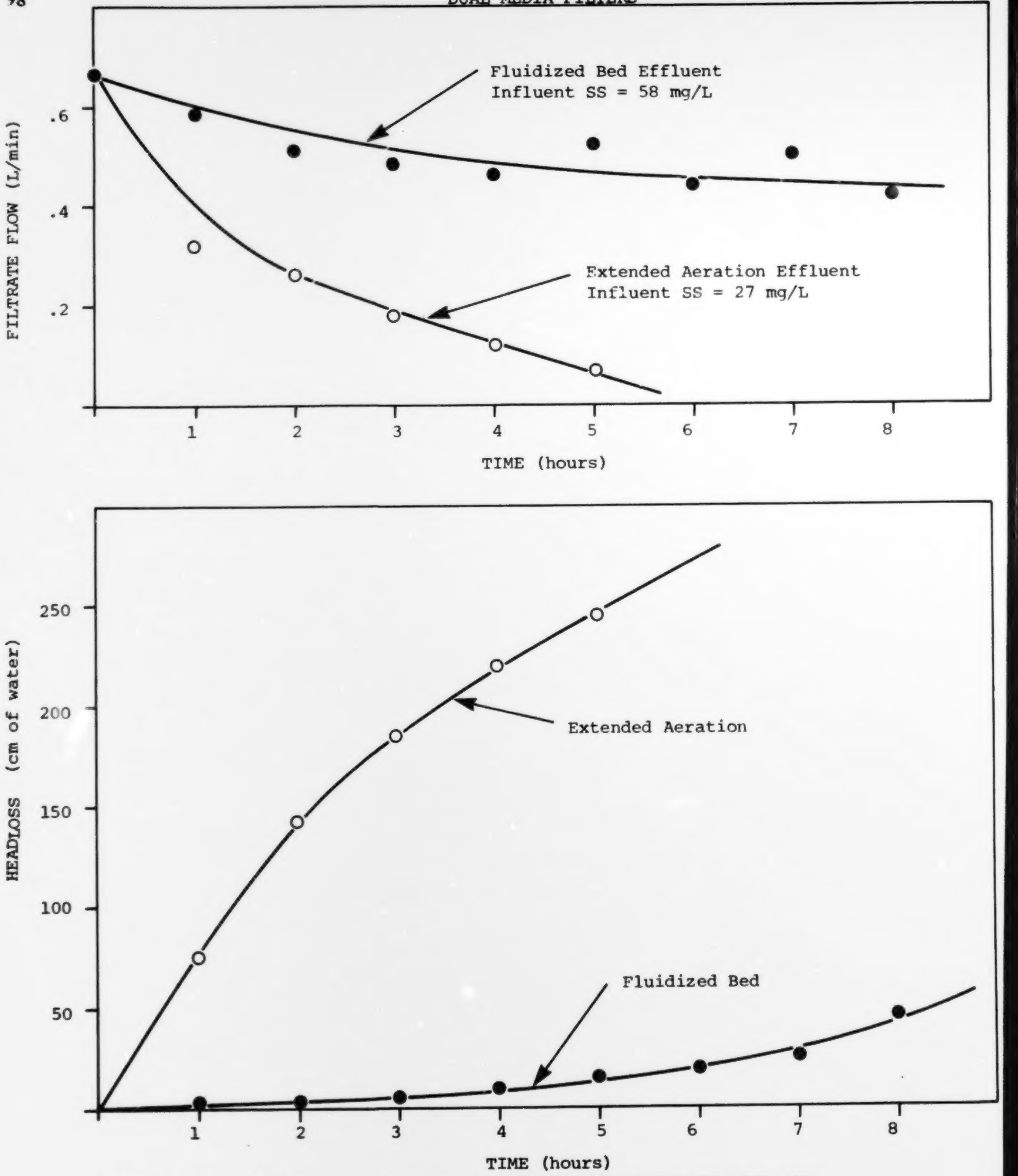


FIGURE 7.13 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS

Hydraulic Loading = 81.4 L/min•m2

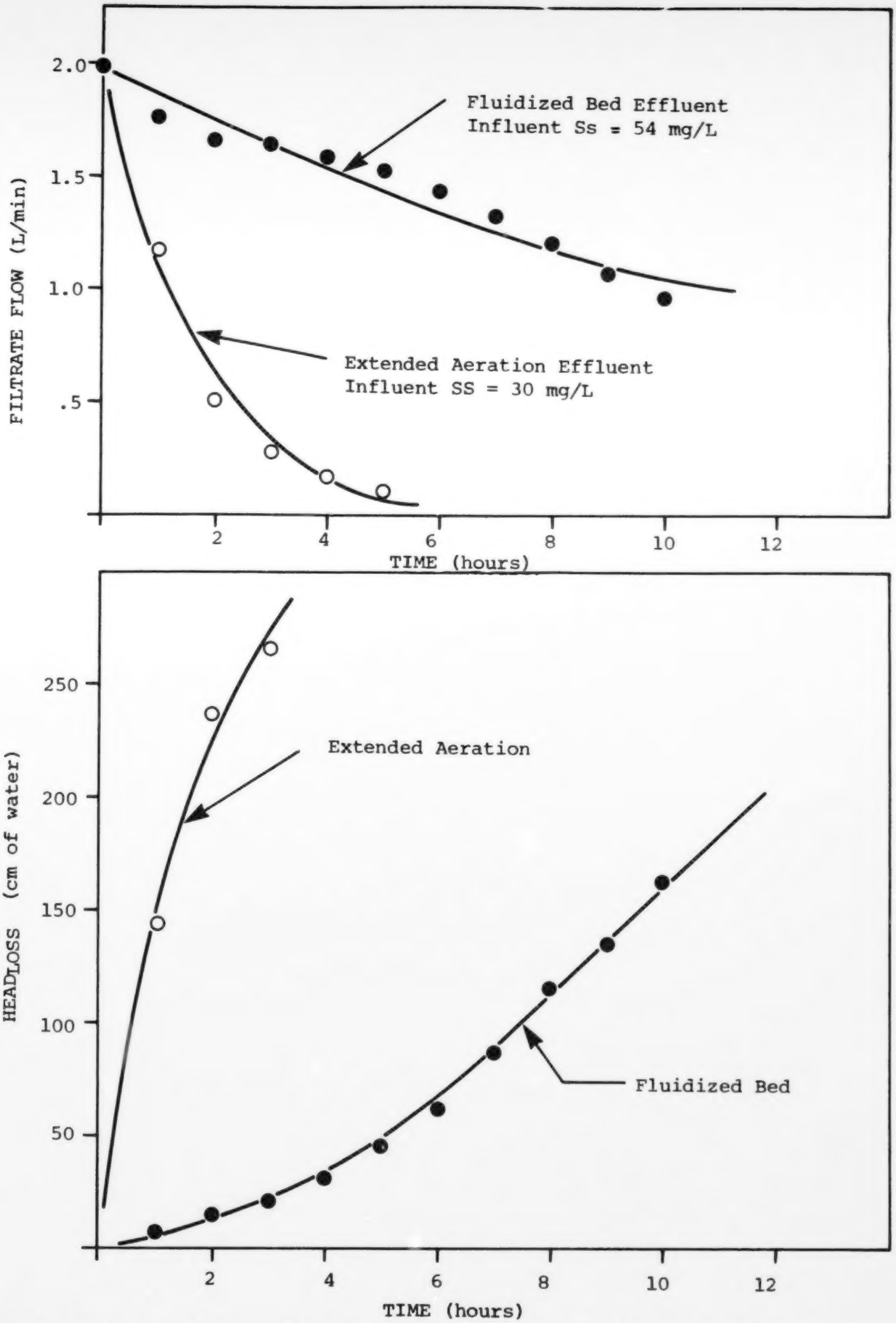


FIGURE 7.14 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS
Hydraulic Loading = 244.2 L/min•m2

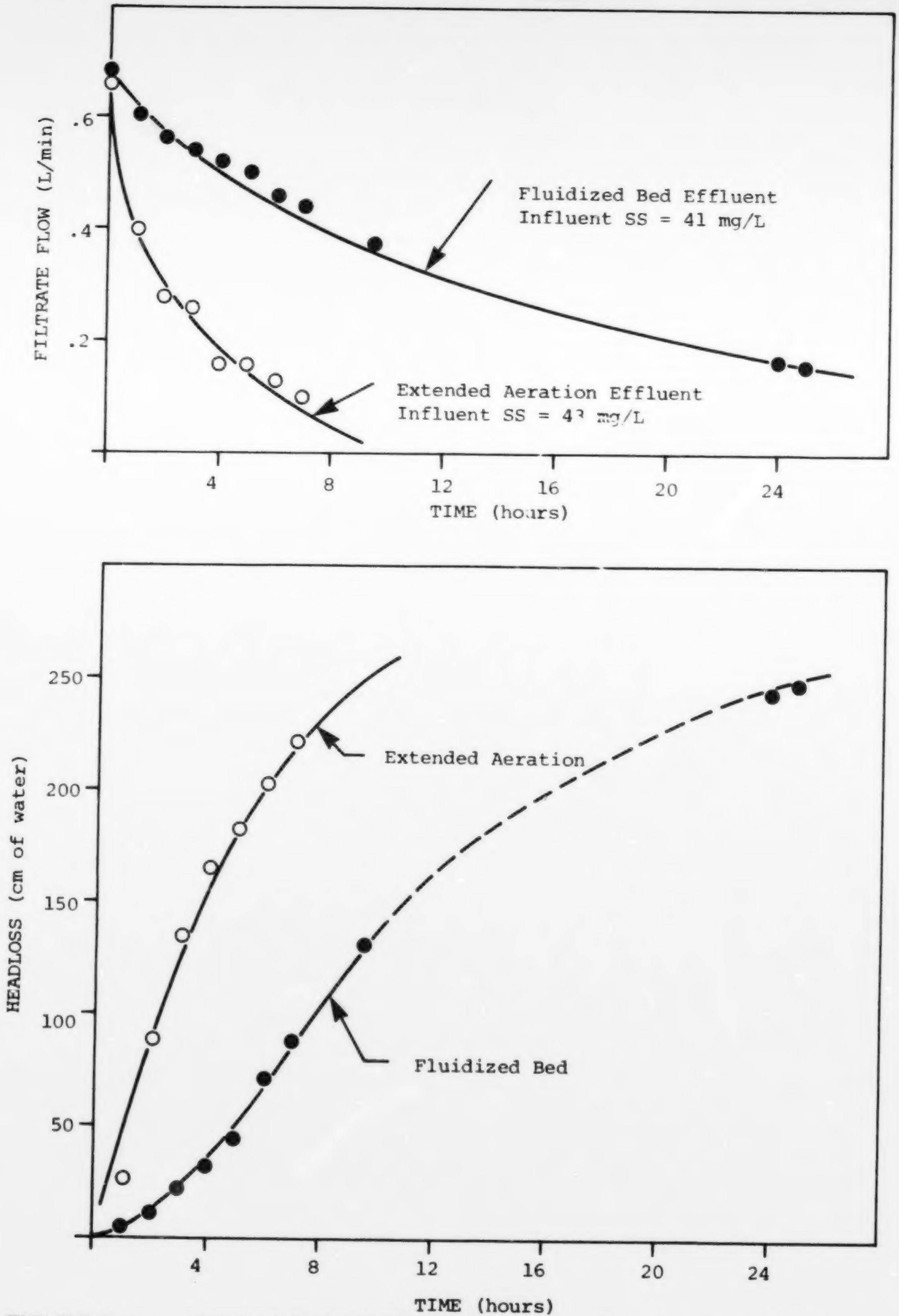


FIGURE 7.15 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS
Hydraulic Loading = 81.4 L/min•m2

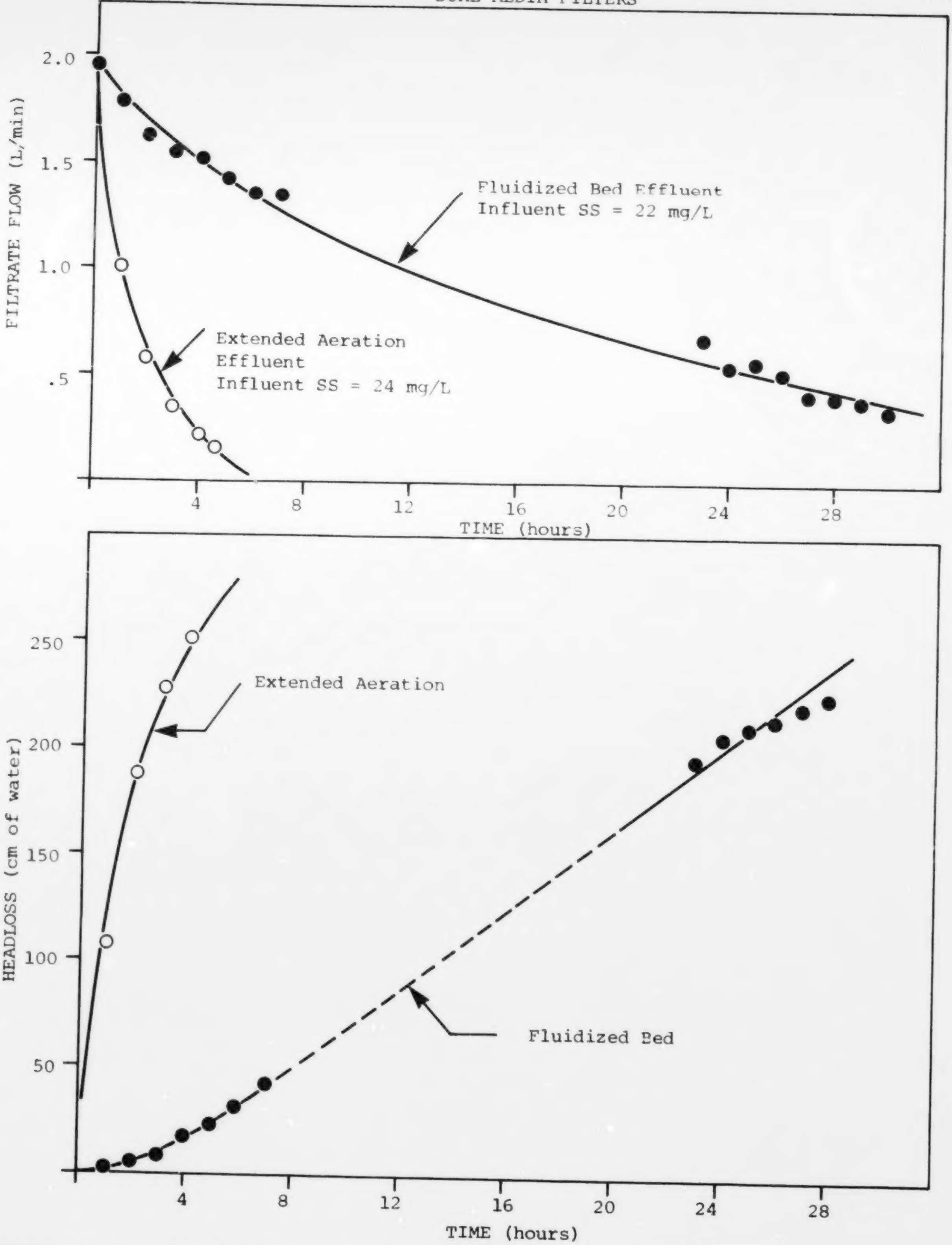


FIGURE 7.16 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS
Hydraulic Loading = 244.2 L/min•m2

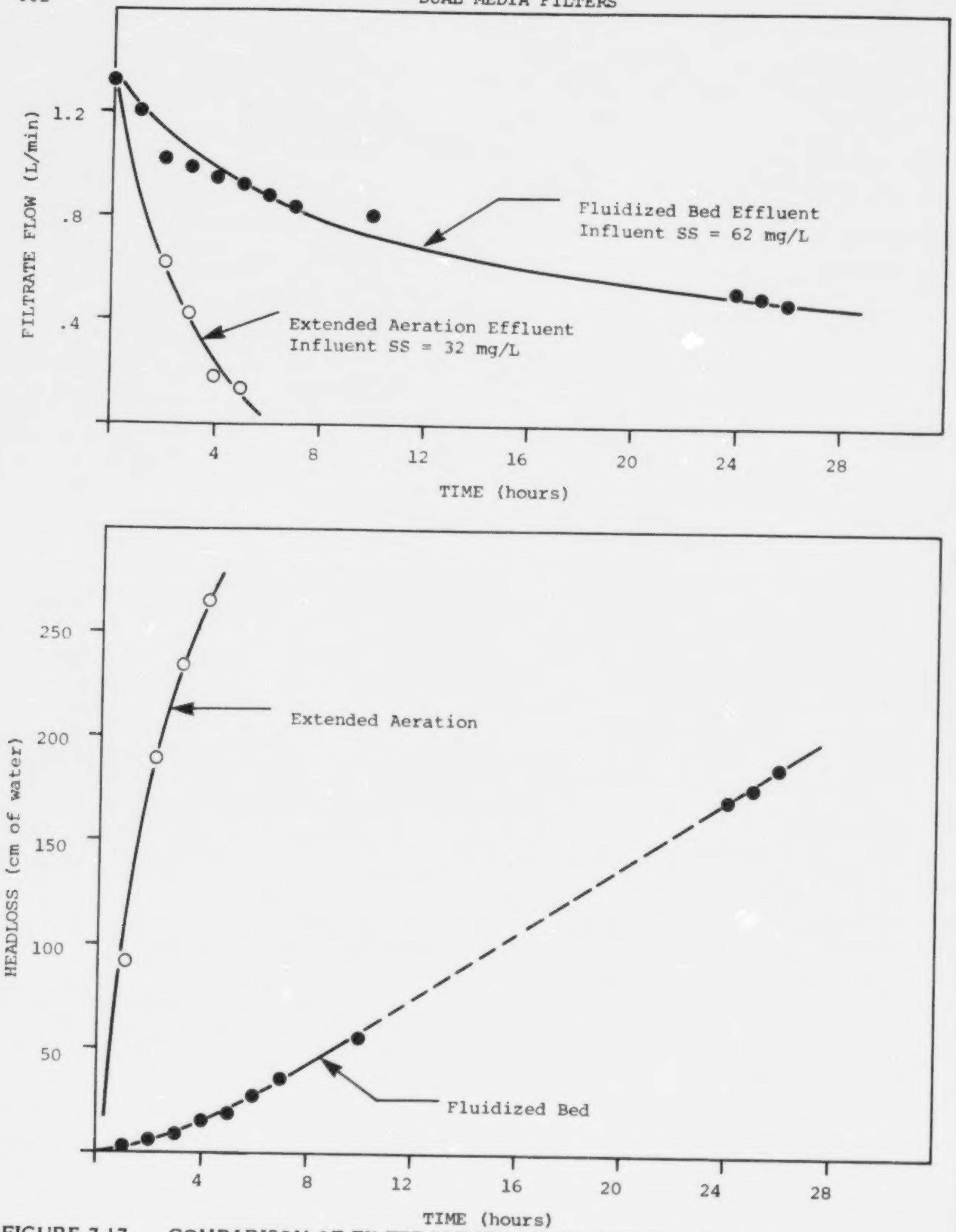


FIGURE 7.17 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS
Hydraulic Loading = 162.8 L/min•m^2

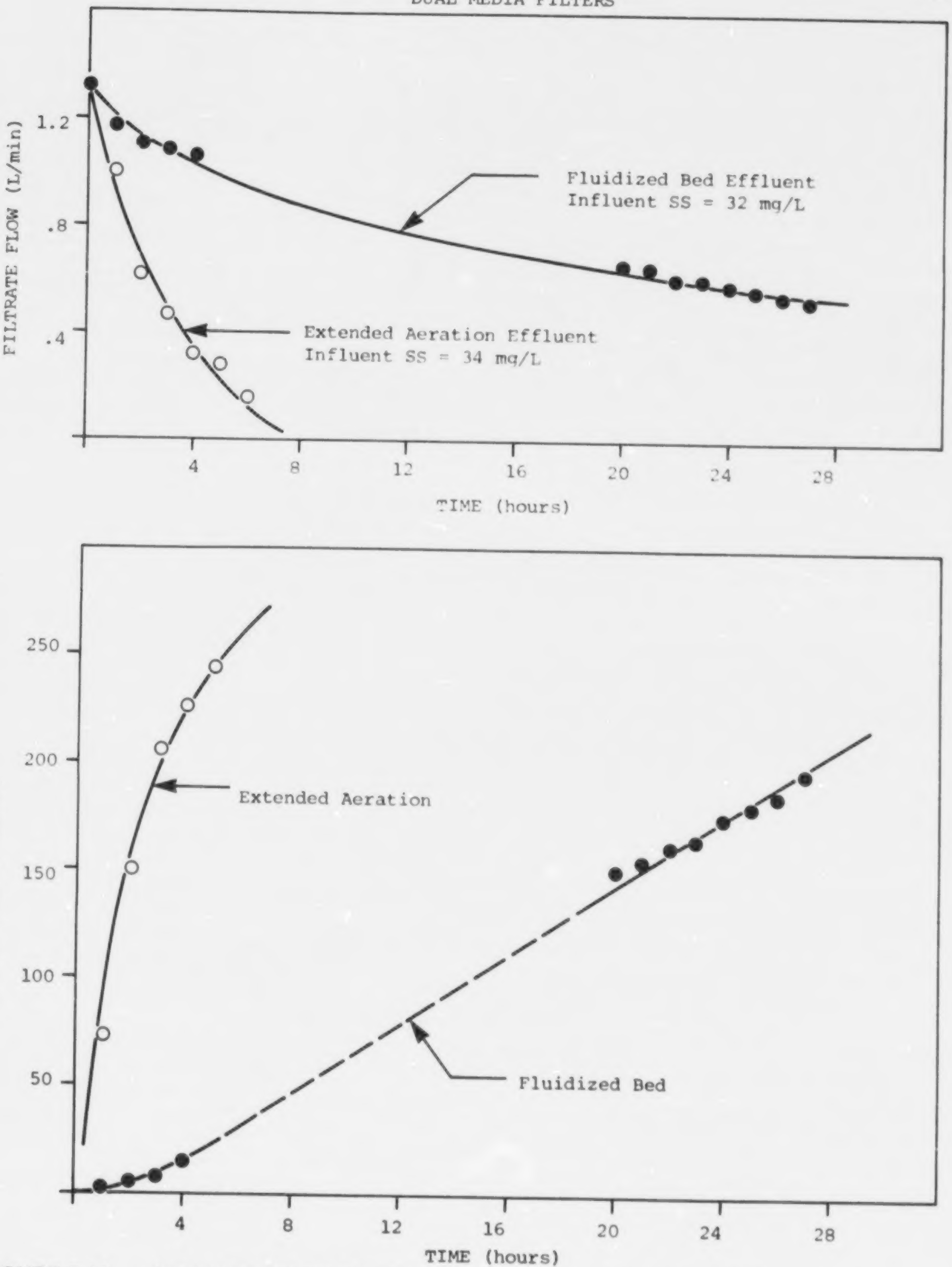


FIGURE 7.18 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS
Hydraulic Loading = 162.8 L/min•m2

TABLE 7.8 COMPARISON OF FILTERABILITY OF FLUIDIZED BED AND EXTENDED AERATION PLANT EFFLUENTS

Dual Media Filters: 38.1 cm Anthracite (d_{10} = 0.96) and 38.1 cm Sand (d_{10} = 0.70)

Initial Hydraulic Loading (L/min•m^2)	Fluidized Bed Filter							Extended Aeration Plant Filter						
	Influent Quality (mg/L)*			Effluent Quality (mg/L)*			Run time to 200 cm Headloss (hours)	Influent Quality (mg/L)*			Effluent Quality (mg/L)*			Run time to 200 cm Headloss (hours)
	Suspended Solids	BOD$_5$		Suspended Solids	BOD$_5$			Suspended Solids	BOD$_5$		Suspended Solids	BOD$_5$		
		Total	Filtered		Total	Filtered			Total	Filtered		Total	Filtered	
81.4	58	35	12	10	15	13	13**	27	14	6	5	8	8	3.5
244.2	54	44	20	18	30	20	12	30	11	4	4	4	4	1.5
81.4	41	26	6	6	9	6	16	43	16	4	5	8	7	6.0
244.2	22	17	8	9	15	9	24	24	16	6	1	5	4	2.5
162.8	62	38	13	15	19	13	28	32	16	5	1	3	3	2.5
162.8	32	15	7	13	12	8	27	34	18	3	3	5	4	3.0

* Based on influent and effluent composite samples.

** Estimated based on extrapolation of experimental data.

Filtration of fluidized bed effluent at the lowest hydraulic loading ($81.4\ L/min{\cdot}m^2$) resulted in shorter run times (13 to 16 hours) than at the median hydraulic loading ($162.8\ L/min{\cdot}m^2$). Considerable surface filtration, which was not evident in the filters operated at higher hydraulic loading conditions, was observed at the low hydraulic loading. The depth removal of suspended solids at the higher loadings is responsible for the improved filter run-times. Effluent quality was somewhat better at lower loadings.

At the median hydraulic loading ($162.8\ L/min{\cdot}m^2$), solids removal efficiency on the fluidized bed effluent was 60 to 75 percent. The solids loading did not appear to affect the filter performance significantly. At the highest hydraulic loading ($244.2\ L/min{\cdot}m^2$), removal of suspended solids was 60 to 66 percent. In this case, there was a considerable reduction in filter run-time at increased solids loading.

At all three hydraulic loadings, a filtered effluent containing less than 15 mg/L suspended solids could be attained when the fluidized bed effluent contained approximately 30 mg/L suspended solids. At filter influent suspended solids levels of 50 to 60 mg/L, a filtered effluent (less than 15 mg/L) was achieved only at the lowest condition. The soluble BOD_5 fraction contributed more than two-thirds of the total BOD_5 concentration of the filtered effluent.

Based on the filtration runs, the dual media filter configuration used in Phase II produced an effluent containing 15 mg/L suspended solids or less at a hydraulic loading of $162.8\ L/min{\cdot}m^2$, when the fluidized bed effluent contained suspended solids concentrations of up to 60 mg/L. Similar quality effluent resulted from filter operation at hydraulic loadings of up to $244.2\ L/min{\cdot}m^2$ if the filter influent contained less than about 30 mg/L. Filter operation under these conditions did not significantly reduce the length of the filter run. To attain an effluent BOD_5 concentration of less than 15 mg/L under these conditions required that the fluidized bed process be operated to produce a treated effluent containing less than about 10 mg/L soluble BOD_5.

To attain higher quality filtered effluents (5 mg/L BOD_5 and suspended solids) with the dual media filter configuration used in Phase II required a reduction in the filter hydraulic loading to $81.4\ L/min{\cdot}m^2$ and operation of the fluidized bed to produce an effluent containing less than 40 mg/L suspended solids and less than 5 mg/L soluble BOD_5. The dual media configuration employing finer sand, as in Phase I, could produce such an effluent at a hydraulic loading of $162.8\ L/min{\cdot}m^2$ and similar fluidized bed operating conditions. The filter run length is considerably shortened.

Filter performance might be further improved by optimization of the media used in Phase II for the dual media filter system. Coarser anthracite (d_{10}>0.96 mm) would eliminate the surface blinding problems evident at the low hydraulic loading. Sand of an effective size between that used in Phase I and Phase II (0.48 mm <d_{10} <0.70 mm) would improve the efficiency attained during Phase II while increasing the filter run-times in Phase I. Further pilot-scale testing would be required.

Increasing the suspended solids concentration of the extended aeration plant effluent by adding mixed liquor solids dramatically reduced the filter run-times (Table 7.8). At similar hydraulic loadings and influent suspended solids concentrations, the extended aeration filtration runs were considerably shorter than the fluidized bed filtration runs. Maximum filter run length on the extended aeration plant effluent was six hours using the dual media filter.

Observation indicated that solids removal from the extended aeration plant effluent was due to surface filtration. A blanket of solids rapidly accumulated on the surface of the anthracite, resulting in a rapid increase in headloss across the filter. The filter run length for the extended aeration plant effluent decreased as the filter hydraulic loading was increased.

Effluent quality from the extended aeration plant filter did not appear to be affected by the filter operating parameters. Residual suspended solids levels were consistently less than 5 mg/L. The effluent BOD_5 after filtration was considerably lower than that attainable by filtration of the fluidized bed effluent; the extended aeration plant effluent contained less than 5 mg/L soluble BOD_5.

Backwash of Dual Media Filters: Although no quantitative measurements were made, qualitative observations during backwashing indicated substantial differences in the mechanisms of solids removal and backwash requirements between filters operated on fluidized bed effluents and extended aeration plant effluents.

In dual media filtration of the fluidized bed effluents, except for those conducted at the lowest hydraulic loading, solids accumulation was evident throughout both the sand and anthracite at the end of the filter run. At the low hydraulic loading, most of the solids removed accumulated in the upper half of the anthracite. Little solids accumulated at the filter surface.

The combination of air scouring and backwashing for 10 minutes removed the accumulated solids and returned the filter bed to its original condition. The solids removed by backwashing appeared similar to those entering from the fluidized bed. There was little indication of flocculation.

During filtration of extended aeration plant effluents, solids accumulated in the upper 2.5 to 5 cm of the bed with no visible penetration to the sand layer. Removal of the accumulated solids at the end of the filter run was more difficult. Longer air scouring was necessary to break up the surface cake of solids. Backwashing for up to 40 minutes was necessary under the worst conditions to remove the accumulated solids.

7.3 Summary

Based on the ATP content of fluidized bed influent and effluent suspended solids and reactor biomass, influent solids which were not removed in the biological reactor contributed significantly to the effluent suspended solids concentration. The relative fraction of biomass sloughed from the reactor support media increased as the concentration of suspended solids in the treated effluent decreased.

Effluent suspended solids exhibited relatively poor settling properties. Secondary clarification does not appear to be a feasible means of upgrading fluidized bed effluent to meet a 15 mg/L BOD - 15 mg/L suspended solids discharge requirement. Chemical coagulation with ferric chloride, in conjunction with an anionic polyacrylamide, improved the settling rate and produced a superior effluent. However, further testing would be necessary to determine the feasibility of this approach.

The fluidized bed effluent was readily filtered to a suspended solids content of less than 15 mg/L with a dual media gravity filter containing anthracite (d_{10} = 0.96 mm) and sand (d_{10} = 0.70 mm) at an initial filter hydraulic loading of 162.8 L/min•m^2 when the fluidized bed effluent contained a suspended solids concentration of up to 60 mg/L. A similar quality effluent was produced at an initial filter hydraulic loading of 244.62 L/min•m^2 when the influent contained less than 30 mg/L suspended solids.

To attain a high quality effluent (approximately 5 mg/L BOD_5 and 5 mg/L suspended solids) required filter operation at a low initial hydraulic loadings (81.4 L/min•m^2) and operation of the fluidized bed to produce an effluent containing less than 40 mg/L suspended solids and less than 5 mg/L soluble BOD_5.

8 SEPARATE STAGE NITRIFICATION IN THE BIOLOGICAL FLUIDIZED BED

The biological oxidation of inorganic (ammonia) nitrogen is carried out in the aquatic environment by bacteria of the family Nitrobacteraceae. The two principal genera in biological nitrification are Nitrosomonas, responsible for the oxidation of ammonia to nitrite, and Nitrobacter, responsible for the oxidation of nitrite to nitrate.

Biological nitrification can be achieved in either suspended growth or fixed film biological treatment systems concurrent with carbon oxidation by proper control of organic loading, system SRT, pH, dissolved oxygen concentration and temperature. Alternatively, a separate stage nitrification system can be operated by pretreatment of the wastewater to lower the organic carbon concentration; thus, a higher proportion of nitrifiers is maintained in the nitrification stage, resulting in higher rates of nitrification. According to USEPA(46), an influent BOD_5/TKN ratio of less than 3.0 is characteristic of a separate stage nitrification process, whereas a BOD_5/TKN ratio greater than 5.0 in the wastewater would indicate a combined carbon oxidation-nitrification process.

The fourth phase of this investigation aimed to assess the performance of the biological fluidized bed system operated as a separate stage nitrification reactor. Minimal modifications to the pilot plant allowed operation in the nitrification mode (Section 3.1). The effluent dissolved oxygen concentration was controlled at 5 mg/L.

8.1 Reactor Feed Quality

The source of feed for the fluidized bed reactor was the pilot-scale extended aeration activated sludge plant (Section 7.2.3). This plant was operated to ensure complete nitrification of the raw sewage feed from Burlington Skyway WPCP. Ammonium chloride (CIL Food Grade, BASF, & Allied Chemical Technical Grade) was added to the feed to the fluidized bed reactor at a predetermined, controlled rate. Sodium bicarbonate (Arm & Hammer Brand, Church and Dwight Ltd.) was added at the same time at a rate equivalent to seven to 10 times the ammonia nitrogen addition rate, to control reactor pH and ensure that there was no alkalinity limitation to the nitrification rate.

The extended aeration plant produced an extremely high quality effluent, particularly in terms of the organic carbon concentration. The feed to the fluidized bed reactor consistently contained less than 5 mg/L of filterable BOD_5. The fluidized bed feed BOD_5/TKN ratio was less than 0.4 (Table 8.1).

TABLE 8.1 CHARACTERISTICS OF FLUIDIZED BED FEED DURING NITRIFICATION STUDIES

Parameter	Concentration (mg/L)	
	Mean	Range
FOC	8.5	4 - 30
Total BOD_5	< 8	<5 - 26
Filterable BOD_5	< 5	<5 - 5
Suspended Solids	7	2 - 86
TKN	23.2	2.7 - 50.1
NH_3 - N	19.9	0.9 - 49.3
NO_3 - N	14.2	2.4 - 24.2
NO_2 - N	0.4	0.1 - 6.2
Alkalinity (as $CaCO_3$)	344	200 - 550
pH	-	7.0 - 8.2
$\frac{BOD_5}{TKN}$ Ratio	<0.34	-

Due to the complete nitrification of the raw wastewater in the extended aeration plant, the fluidized bed reactor feed contained up to 25 mg/L nitrate-N. However, this concentration of nitrate-N would not inhibit nitrification in the fluidized bed(47). There was no indication of nutrient limitations despite the complete biological pretreatment of the wastewater.

8.2 Start-up and Acclimation

After the carbon oxidation phase, the fluidized bed system, which contained approximately 9.5 g/L of a heterotrophic biomass, was acclimated to nitrification conditions by increasing the system HRT to approximately 40 minutes while maintaining primary clarified wastewater as the system feed. At the resultant organic loading of approximately 0.26 g BOD_5/g BVS•d and SRT_a of six days (T = 21°C), a small degree of nitrification was evident after approximately 12 days based on a net production of nitrite-N in the treated effluent. The system HRT was increased to 52 minutes, decreasing the process organic loading to approximately 0.22 g BOD_5/g BVS•d. After four more days of operation, there was no significant increase in the nitrification efficiency. Therefore, extended aeration plant effluent with ammonium chloride addition was commenced as feed on October 20, 1979. The system HRT was maintained at 52 minutes (based on the empty bed volume). There was an immediate improvement in the rate of nitrite production followed by conversion of nitrite to nitrate after about 10 days when the Nitrobacter population had become established.

With the substitution of low BOD_5 extended aeration plant effluent for primary clarified raw sewage, the volatile solids content of the biological reactor decreased steadily as a nitrifying bacterial population replaced the established heterotrophic population. The loss of solids from the system with a parallel reduction in expanded bed height continued for approximately 40 days after the initiation of extended aeration feed. At this time, the bed volatile solids concentration had stabilized at approximately 2.0 g/L and the expanded bed height had been reduced to about 3 metres from 3.9 metres despite an increase in total flux from 0.46 m^3/m^2•min to 0.71 m^3/m^2•min. The change in bed conditions during acclimation is shown in Figure 8.1. The total biological solids in the reactor decreased from approximately 2500 g at the start of extended aeration feed to about 400 g after 40 days. The hydraulic flux to the reactor was increased to 0.81 m^3/m^2•min after about 60 days of operation to expand the bed back to the top of the reactor. However, there was no significant increase in the reactor volatile solids concentration until after 110 days.

The nitrification efficiency during acclimation is shown in Figure 8.2. A gradual improvement in the system nitrification over the first 30 days after initiation of extended aeration feed was evident and the rate appeared to stabilize at approximately 0.23 g NH_3-N oxidized/g BVS•d at a temperature of 20°C. However, the bed volatile solids concentration was still declining. To increase the growth rate of nitrifiers, the temperature was increased to 25°C. The biological system responded with an increase in nitrification rate reaching a maximum of approximately 1.0 g NH_3-N oxidized/g BVS•d. The bed volatile solids concentration had stabilized at 2.2 g/L and, by increasing the flux to 0.81 m^3/m^2•min, the bed was expanded to the wasting valve at 3.9 m.

On December 23 (Day 68), failure of the extended aeration plant return sludge pump resulted in a high concentration of mixed liquor solids in the feed to the fluidized bed system. From December 27 to January 3 (Days 72 to 79), problems with the ammonium chloride feed pump reduced the influent ammonia-nitrogen concentration to an

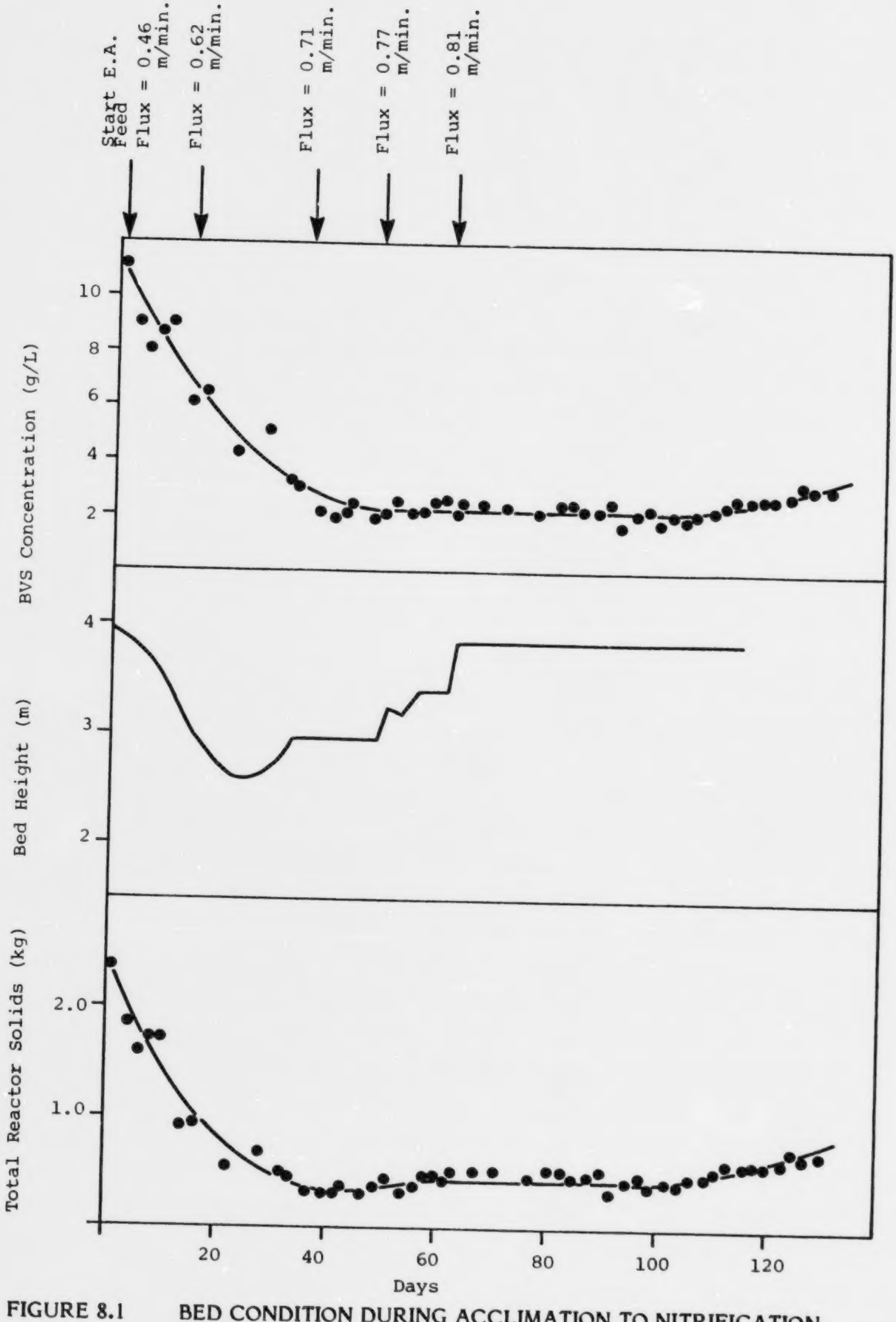


FIGURE 8.1 BED CONDITION DURING ACCLIMATION TO NITRIFICATION CONDITIONS

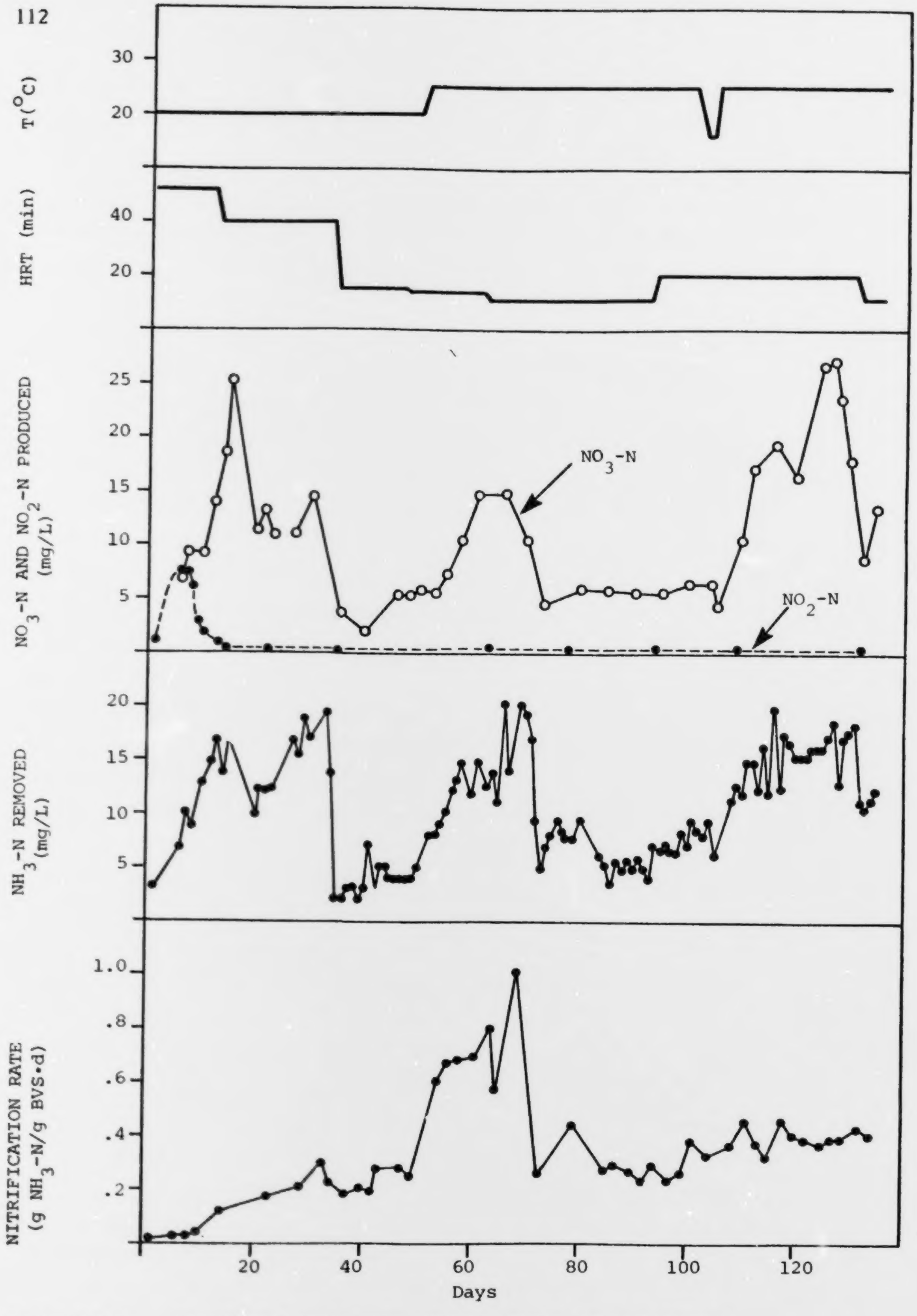


FIGURE 8.2 NITRIFICATION EFFICIENCY DURING ACCLIMATION

average level of 14.0 mg/L NH_3-N from the previous level of 20 to 25 mg/L and, for short periods, interrupted completely the addition of ammonium chloride. In addition, on December 27 (Day 72), technical grade ammonium chloride (Allied Chemical) was substituted for the food grade ammonium chloride (CIL) used exclusively during the previous acclimation. The nitrification rate dropped dramatically after Day 70 to approximately 0.3 g NH_3-N/g BVS•d.

Comparison of the rate of nitrification in batch tests using extended aeration plant mixed liquor and fluidized bed biomass showed that the type of ammonium chloride had no significant effect on the ammonia oxidation rate. Although the interruption in ammonia chloride feed and the upset of the extended aeration plant may have affected the nitrification rate, continued operation of the fluidized bed at 25°C for a further 60 days did not result in nitrification rates approaching previous levels. The nitrification rate appeared to reach a steady state level of approximately 0.4 g NH_3-N/g BVS•d.

It would appear that the extremely high nitrification rates observed between Day 55 and Day 70 were indicative of an unstable biomass condition in terms of the population dynamics. Based on theoretical considerations, the nitrification rate data shown in Figure 8.3 have been developed to show the effect of the fraction of nitrifiers (f) on the ammonia oxidation rate(46). From these data, a nitrification rate of approximately 1.0 g NH_3-N/g VSS•d is achievable at 25°C if the fraction of nitrifiers approaches approximately 0.35. On a similar basis, the data in Table 8.2 indicate that a nitrifier fraction of 0.35 would be attained at a BOD_5/TKN ratio of 0.5, similar to the ratio in the extended aeration plant feed. However, several authors have shown that it is difficult to maintain nitrification in a suspended growth system at low BOD_5/TKN ratios as the biomass is poorly flocculated(47,48). A similar situation may exist in a fixed film process, although there is no information available. The high fraction of nitrifiers necessary to attain these rates may represent an unstable biofilm condition. The fact that nitrification rates approaching 1.0 g NH_3-N/g BVS•d could not be re-established despite prolonged operation appears to support this, but additional investigation would be necessary.

Based on the acclimation period data, it was evident than an extended operation was necessary to establish steady state conditions in the fluidized bed under separate stage nitrification. In fact, absolute steady state conditions were never attained as the bed volatile solids concentration continued to increase, reaching a maximum level of approximately 4.7 g/L on March 16 (Day 153 after initiation of extended aeration feed).

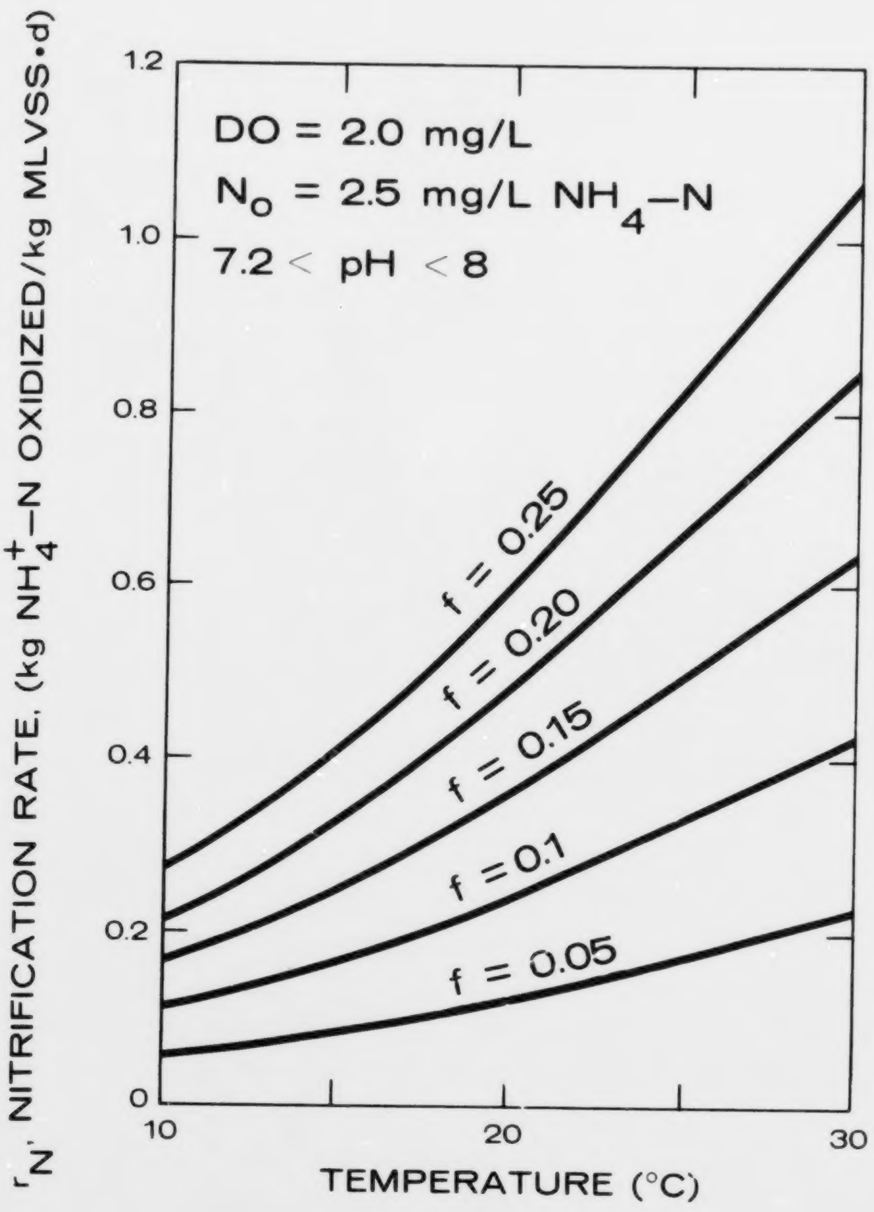


FIGURE 8.3 EFFECT OF TEMPERATURE AND FRACTION OF NITRIFIERS ON NITRIFICATION RATE*

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*From USEPA(46)

TABLE 8.2 RELATIONSHIP BETWEEN NITRIFIER FRACTION AND THE BOD_5/TKN RATIO*

BOD_5/TKN Ratio	Nitrifier fraction	BOD_5/TKN Ratio	Nitrifier fraction
0.5	0.35	5	0.054
1	0.21	6	0.043
2	0.12	7	0.037
3	0.083	8	0.033
4	0.064	9	0.029

*From USEPA(46).

Due to the time required to attain steady state conditions after a significant change in the operating parameters, an intensive steady-state experimental design, such as utilized for the carbon oxidation phase of the investigation (Section 4.1.1), was not practical. A limited number of pseudo-steady state runs were conducted at constant temperature (25°C). System hydraulic retention time (HRT) and ammonia-nitrogen loading rate were the principal variables manipulated. Except in Run 5, the recycle ratio (Q_R/Q_I) was varied to maintain a constant hydraulic flux (0.81 m^3/m^2•min) so that constant bed expansion was maintained. Operating data collected after Day 94 (Figure 8.2) were considered to represent pseudo-steady state conditions.

8.3 Pseudo-Steady State Results

The results of the five pseudo-steady state runs are shown in Figures 8.4, 8.5 and 8.6. Pseudo-steady state conditions were considered established when the ammonia oxidation rate had stabilized. During Run 2, the ammonium chloride feed system failed. After the ammonium chloride addition was re-established, the nitrification rate returned to the pre-upset level after about 10 days. However, performance of the process was erratic. Experimental Runs 3 and 4 were conducted at identical retention times and hydraulic fluxes with the feed concentration reduced to 8.6 mg/L NH_3-N in Run 4 compared to 26.5 mg/L NH_3-N in Run 3.

In Run 5, the total hydraulic flux was reduced to 0.31 m^3/m^2•min to evaluate the performance at similar ammonia-nitrogen loadings to Run 4 but at a lower recycle ratio (Q_R/Q_I = 3.6). The recycle ratio was decreased gradually from 11.0 to 3.6 over five

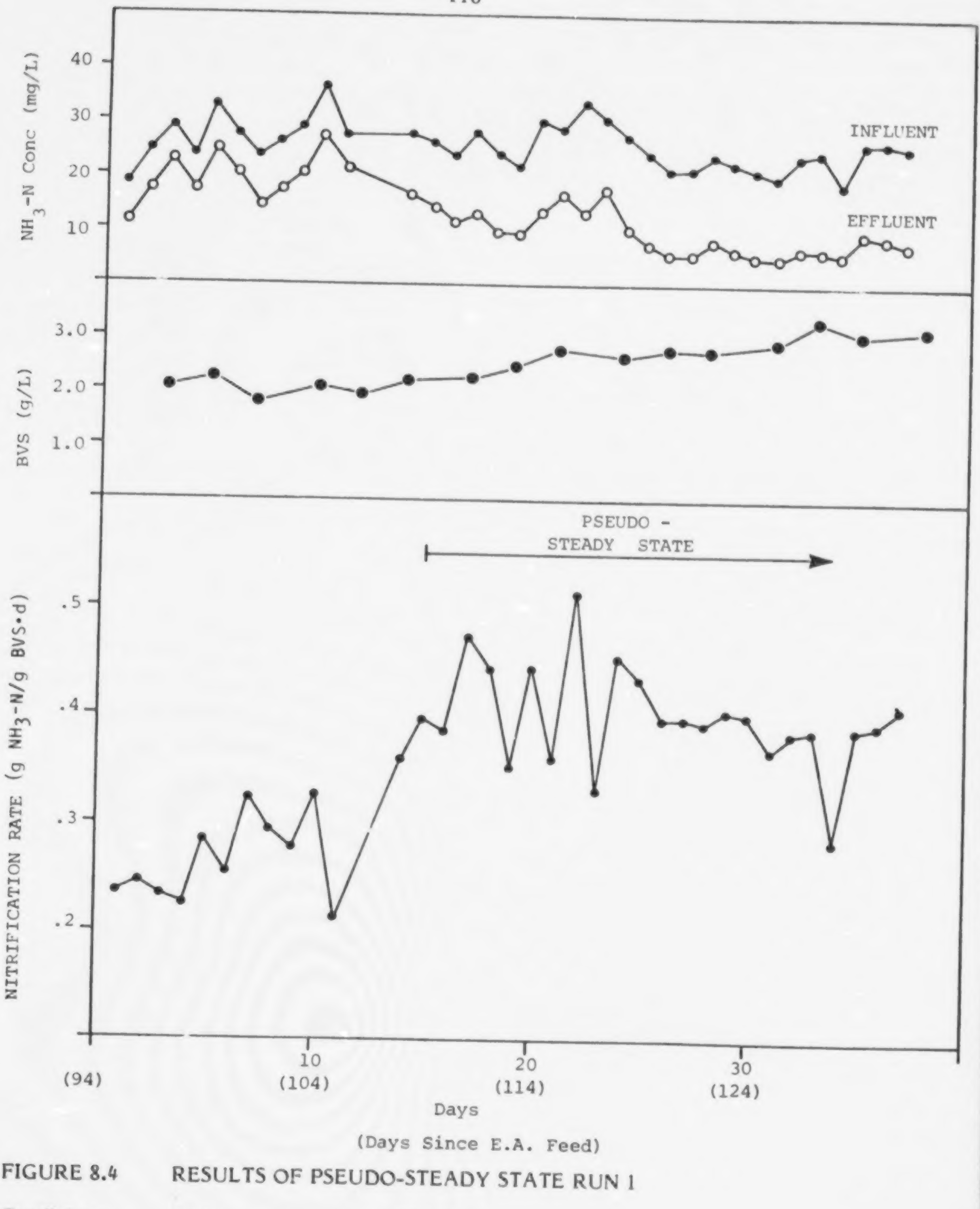

FIGURE 8.4 RESULTS OF PSEUDO-STEADY STATE RUN 1

Conditions:

HRT = 20 min.
RECYCLE = 5.1
TEMP = 25°C
FLUX = 0.81 $m^3/m^2 \cdot min.$

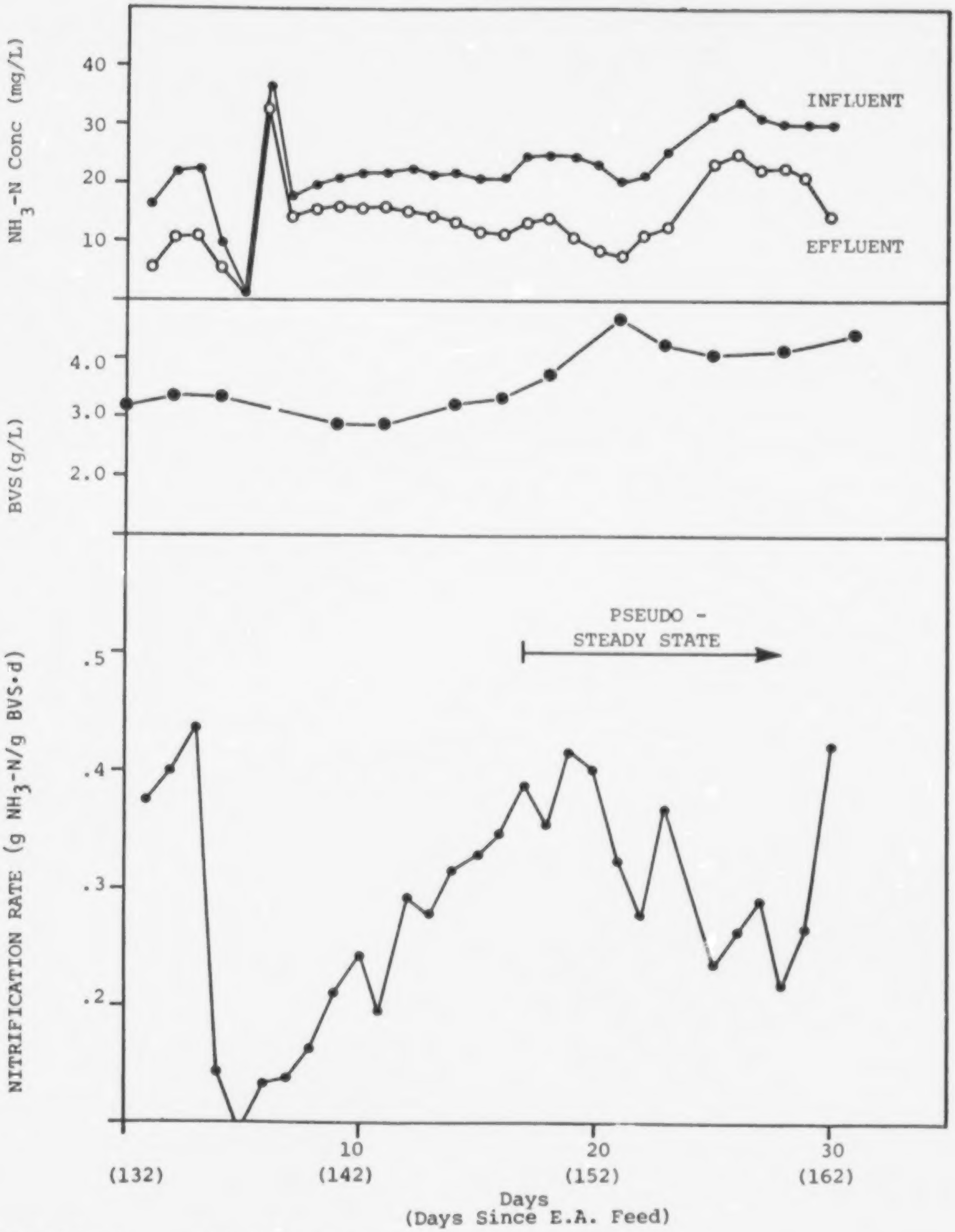

FIGURE 8.5 RESULTS OF PSEUDO-STEADY STATE RUN 2

Conditions:

HRT = 12 min.
RECYCLE = 2.5
TEMP = 25°C
FLUX = 0.81 m^3/m^2•min.

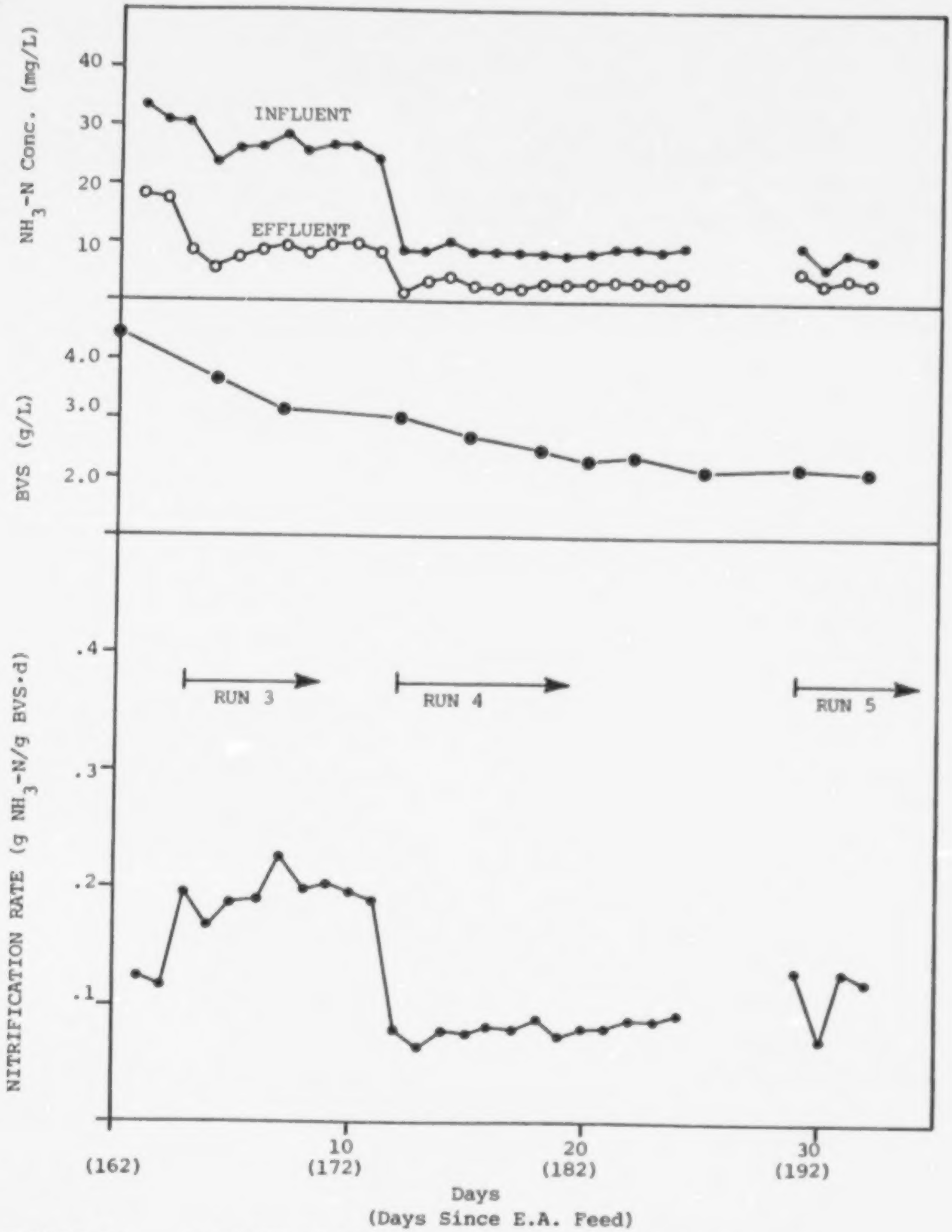

FIGURE 8.6 RESULTS OF PSEUDO-STEADY STATE RUNS 3, 4 AND 5

		Runs 3, 4	Run 5
Conditions:	HRT (min)	40.	26
	RECYCLE	11.	3.6
	TEMP (°C)	25	25
	FLUX ($m^3/m^2 \cdot min$)	0.81	0.31

days to reduce the impact of the lower hydraulic flux. The bed height at these flux conditions was 2.7 m, equivalent to an empty bed volume of 136 L. The nominal HRT, based on this empty bed volume and the extended aeration feed rate, was 26 minutes. Operation of the system at these flux conditions resulted in poor fluidization of the sand and difficulties in maintaining control of the biofilm thicknesses in the top of the bed (Section 4.1.2). The increased turbulence resulted in a loss of about 160 g of biomass from the system while the hydraulic flux was being reduced. Due to the operating difficulties under these conditions, only four days of data could be collected during Run 5 without a proper acclimation period. Although steady state operating conditions were not likely attained, the data appear to agree with the results collected previously.

The performance of the fluidized bed system during the pseudo-steady state runs is summarized in Table 8.3. The nitrification rates in Runs 1 and 2 are comparable to rates observed in full-scale separate sludge air activated sludge systems at similar pH conditions and low BOD_5/TKN ratios(46). For example, the peak nitrification rate observed at the Blue Plains plant (air activated sludge, pH 6.8 to 7.2, BOD_5/TKN = 1.3, 25°C) was approximately 0.25 to 0.35 g NH_3-N/g MLVSS•d(46). There was no significant difference between the nitrification rates observed in Runs 1 and 2, despite the erratic performance during Run 2.

During Run 3, the hydraulic retention time was increased to 40 minutes, reducing the ammonia-nitrogen loading to approximately 0.3 g NH_3-N/g BVS•d. Based on Runs 1 and 2, this loading condition should have resulted in essentially complete removal of ammonia-nitrogen from the secondary effluent. However, the average effluent NH_3-N concentration during Run 3 was 8.4 mg/L. The rate of nitrification was reduced to 0.20 g NH_3-N/g BVS•d. Similarly during Run 4, when the ammonia-nitrogen loading was reduced to about 0.12 g NH_3-N/g BVS•d by decreasing the influent NH_3-N concentration to 8.6 mg/L, the biological nitrification rate was 0.08 g NH_3-N/g BVS•d and the fluidized bed effluent contained an average of 3.1 mg/L NH_3-N.

The effect on process ammonia-N loading on the ammonia removal rate is shown in Figure 8.7. At high loadings (>0.55 g NH_3-N/g BVS•d), the nitrification rate was relatively constant at approximately 0.4 g NH_3-N oxidized/g BVS•d. This represents the maximum nitrification rate at these conditions. At lower loading conditions, the removal efficiency was relatively constant at 60 to 70 percent. The form of the relationship between loading and removal (Figure 8.7) is identical to that identified by Odegaard and Rusten(49) for nitrification in a rotating biological contactor. The authors reported a

TABLE 8.3 SUMMARY OF PSEUDO-STEADY STATE NITRIFICATION RUNS

	Pseudo-Steady State Run No.				
	1	2	3	4	5
Operating Conditions					
HRT (min)*	20.	11.7	40	40	26
Recycle Ratio	5.1	2.5	11.0	11.0	3.6
Hydraulic Flux ($m^3/m^2 \cdot min$)	0.81	0.81	0.81	0.81	0.31
Temperature (°C)	25	25	25	25	25
Effluent DO (mg/L)	5	5	5	5	5
Avg. BVS (g/L)	2.78	4.21	3.25	2.56	2.06
Average Influent Quality (mg/L)					
FOC	8.	7.	8.	7.	10.
TKN	26.7	29.1	31.4	11.2	12.6
NH_3-N	24.8	27.1	26.5	8.6	8.2
NO_3-N	16.4	14.1	10.8	13.0	11.4
NO_2-N	0.4	0.2	0.2	0.3	0.2
Suspended Solids	6.	7.	7.	3.	4.
pH (Range)	7.7 - 8.3	7.7 - 8.2	7.8 - 8.1	7.6 - 8.1	7.8
Average Effluent Quality (mg/L)					
FOC	7.	6.	7.	6.	8.
TKN	10.7	17.7	11.0	5.5	7.6
NH_3-N	9.4	16.0	8.4	3.1	4.0
NO_3-N	35.3	24.2	28.4	18.8	14.8
NO_2-N	0.5	0.8	0.5	0.5	0.3
Suspended Solids	4.	6.	7.	5.	5.
pH (Range)	6.6 - 7.4	6.7 - 7.3	6.8 - 7.1	7.1 - 7.3	7.1
Average Nitrification Rate (g NH_3-N/g BVS•d)	0.39	0.33	0.20	0.08	0.11

*HRT based on extended aeration feed rate and empty bed volume.

critical loading of about 100 to 150 mg NH_4-N/m^2•h over which the nitrification rate was at its maximum and independent of the loading and below which, the nitrification rate was linearly dependent on the nitrification load.

Odegaard and Rusten(49) also demonstrated that, for nitrification in an RBC, zero-order kinetics governed the process at concentrations higher than 4 to 6 mg NH_4-N/L. Below this concentration, the nitrification rate was concentration dependent. The authors assumed a first-order concentration dependency below the limiting concentration.

The pseudo-steady state fluidized bed nitrification rate data also appeared to exhibit a concentration dependency. Figure 8.8 illustrates the effect of the reactor influent ammonia-N concentration on the nitrification rate. In this case, the reactor influent concentration was calculated based on the ammonia-N concentration in the extended aeration plant feed and the dilution of the reactor influent stream by the effluent recycle. Considerable scatter is evident in these data. As the internal mixing regime within the fluidized bed is plug flow in nature, there is a decrease in ammonia-nitrogen concentration over the height of the reactor. This may be partially responsible for the variability in the data. Furthermore, most of the high rate data was collected during Runs 1 and 2 at a relatively low recycle ratio, compared to the lower rate data collected at a high recycle ratio in Runs 3 and 4. Despite the variability in the data, the limiting NH_3-N concentration above which the removal efficiency was independent of the influent concentration appears to be 5 to 10 mg/L NH_3-N.

The effect of ammonia-nitrogen loading on the treated effluent ammonia concentration is shown in Figure 8.9. Throughout the pseudo-steady state runs it was difficult to produce an effluent containing less than 1.0 mg/L NH_3-N. The minimum concentration achieved was 1.7 mg/L NH_3-N during Run 4.

8.4 Nitrification Kinetics in the Biological Fluidized Bed Process

As the pseudo-steady state performance data suggested a concentration dependency in the nitrification rate, a series of batch kinetics evaluations were conducted on the reactor. The fluidized bed was operated under complete recycle conditions and the rate of ammonia-nitrogen oxidation was monitored after the introduction of a spike of ammonium chloride. The system pH was controlled by the manual addition of sodium bicarbonate. The dissolved oxygen concentration at the top of the fluidized bed reactor was controlled at about 5 mg/L. The hydraulic flux on the system during the batch tests was the same as that during continuous operation (0.81 m^3/m^2•min).

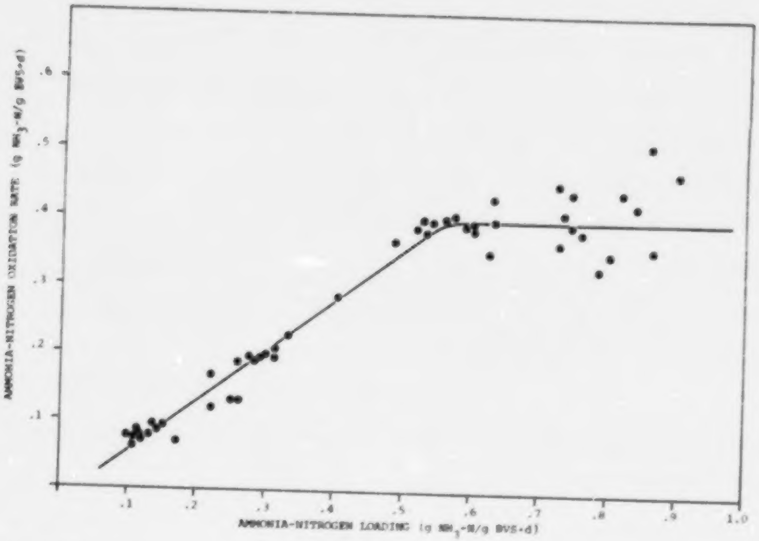

FIGURE 8.7 EFFECT OF AMMONIA LOADING ON NITRIFICATION RATE

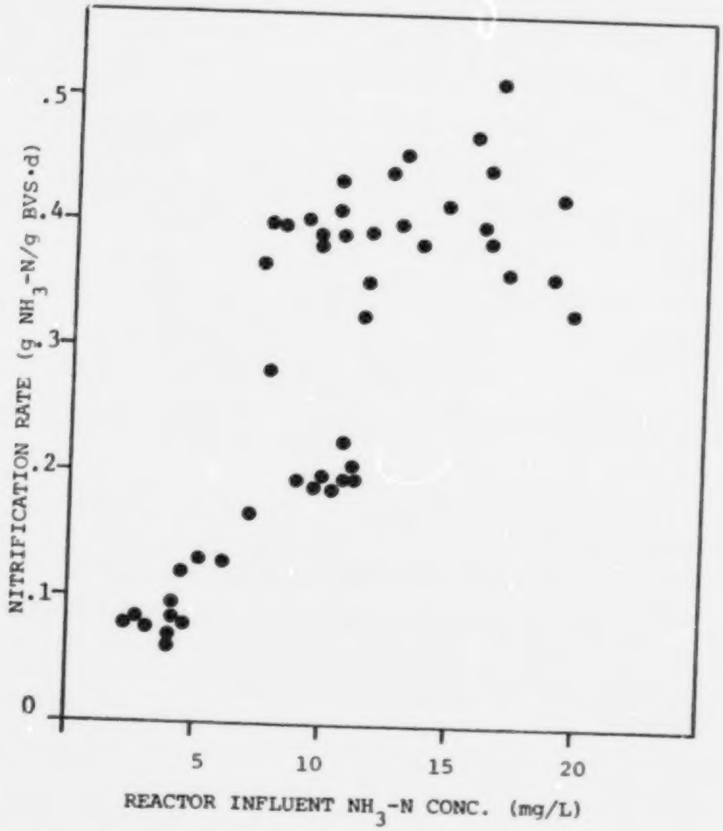

FIGURE 8.8 EFFECT OF REACTOR INFLUENT AMMONIA-N CONCENTRATION ON NITRIFICATION RATE

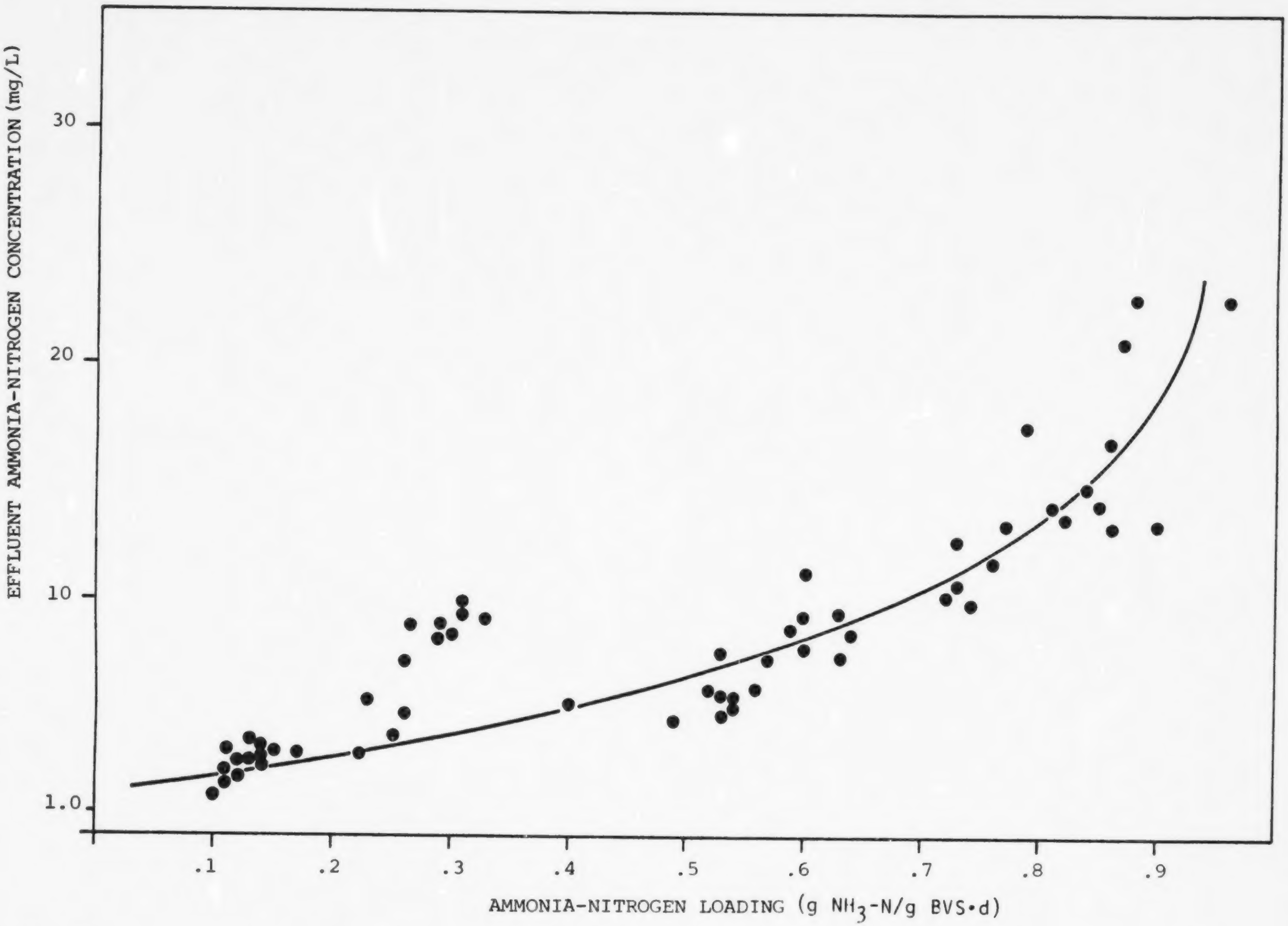

FIGURE 8.9 EFFECT OF AMMONIA LOADING ON FLUIDIZED BED EFFLUENT QUALITY

The results of four batch tests are shown graphically in Figures 8.10 to 8.13. In all cases, the non-linear nature of the process kinetics is evident. A non-linear least squares estimation technique was used to fit the experimental data from the batch runs to zero-, half-and first-order kinetic models where:

$$-\frac{dC}{dt} = kC^n \; ; n = 0, 1/2, 1 \qquad (8.1)$$

The model predictions resulting from the best-fit models are shown in Figures 8.10 to 8.13.

The method of model discrimination by likelihoods(50) was used to confirm the apparent good fit of the bulk half-order nitrification kinetics. The results of the model discrimination test for the four batch kinetic runs are shown in Table 8.4. In all cases, the half-order model was preferred. In experimental runs 1, 2 and 3, the half-order model was strongly preferred. According to Reilly(50), a likelihood ratio of 10 between pairs of models suggests a difference in model plausibility, while a ratio of 100 denotes a strong preference for the favored model.

TABLE 8.4 RESULTS OF MODEL DISCRIMINATION FOR BATCH NITRIFICATION RUNS

Date	Run No.	Model Order	Residual Sum of Squares	Model Likelihood after Reilly(50)
March 13/80	1	Zero	175.5	1.0
		Half	7.9	2.5×10^{36}
		First	101.9	9.6×10^{15}
March 20/80	2	Zero	219.0	1.0
		Half	22.5	4.7×10^{42}
		First	193.6	3.3×10^{5}
March 26/80	3	Zero	19.3	2.1×10^{2}
		Half	4.5	3.7×10^{5}
		First	30.1	1.0
March 26/80	4	Zero	45.1	1.0
		Half	1.5	2.9×10^{9}
		First	8.3	9.8×10^{7}

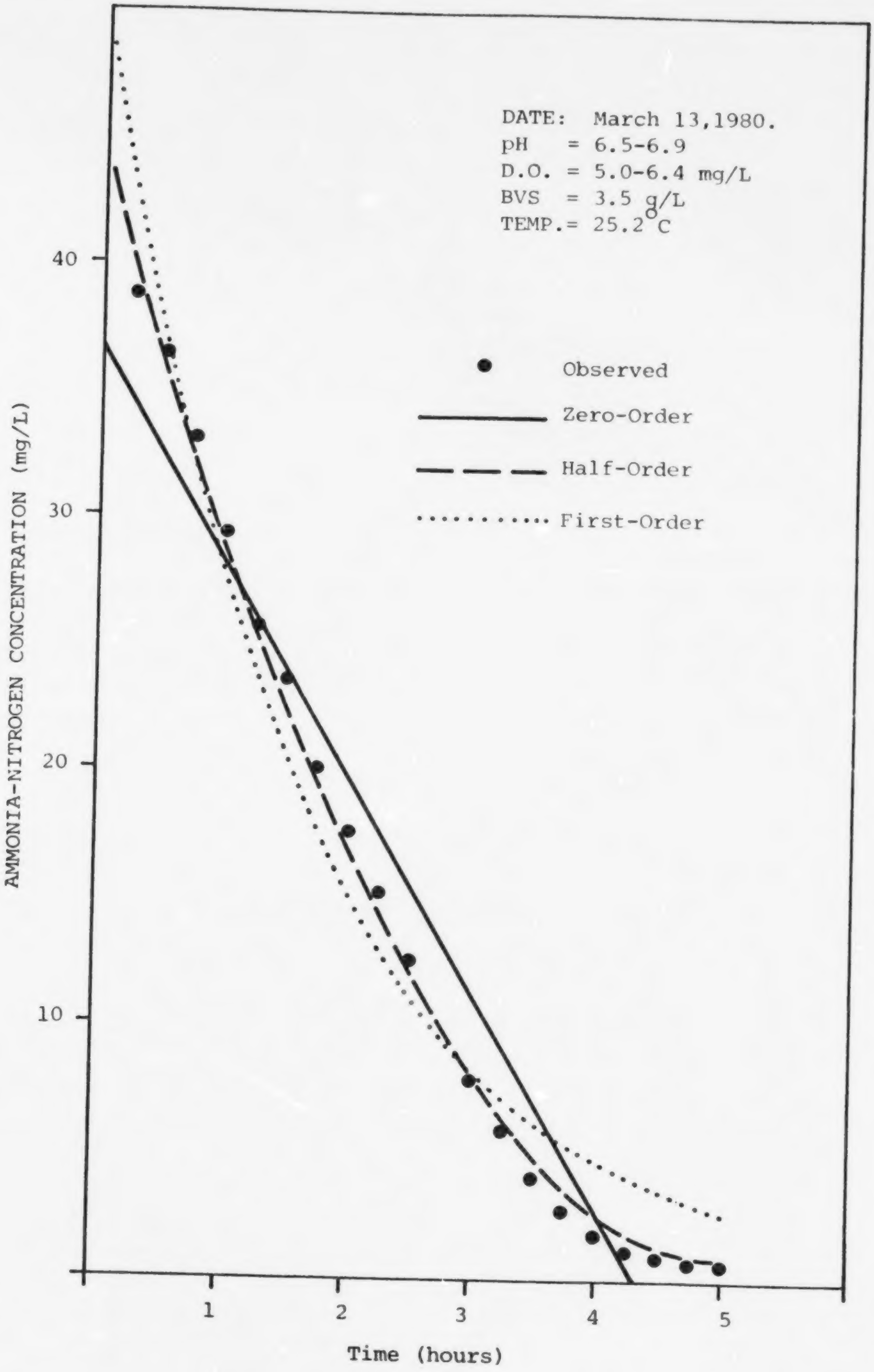

FIGURE 8.10 RESULTS OF FLUIDIZED BED BATCH KINETIC RUN 1

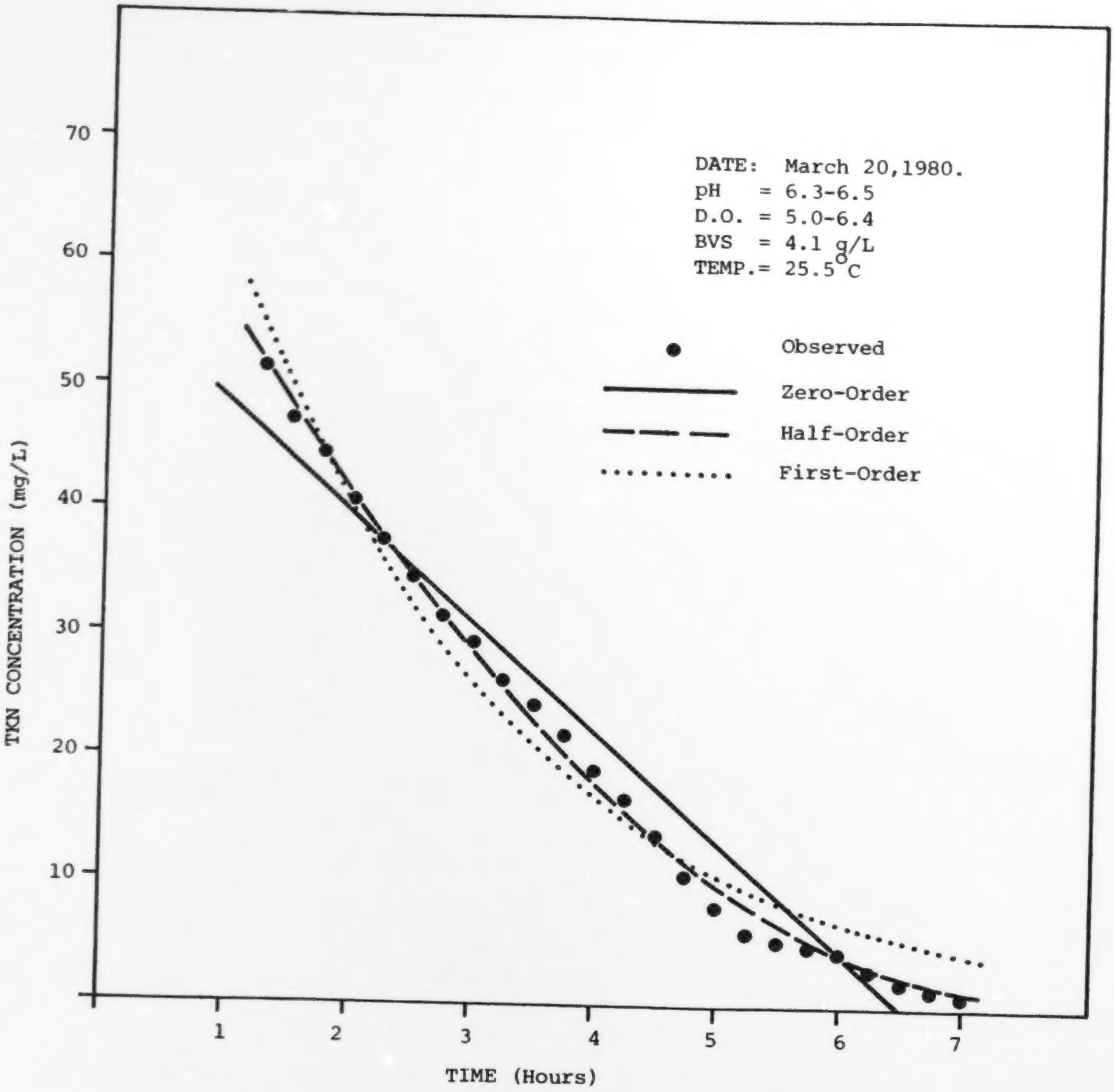

FIGURE 8.11 RESULTS OF FLUIDIZED BED BATCH KINETIC RUN 2

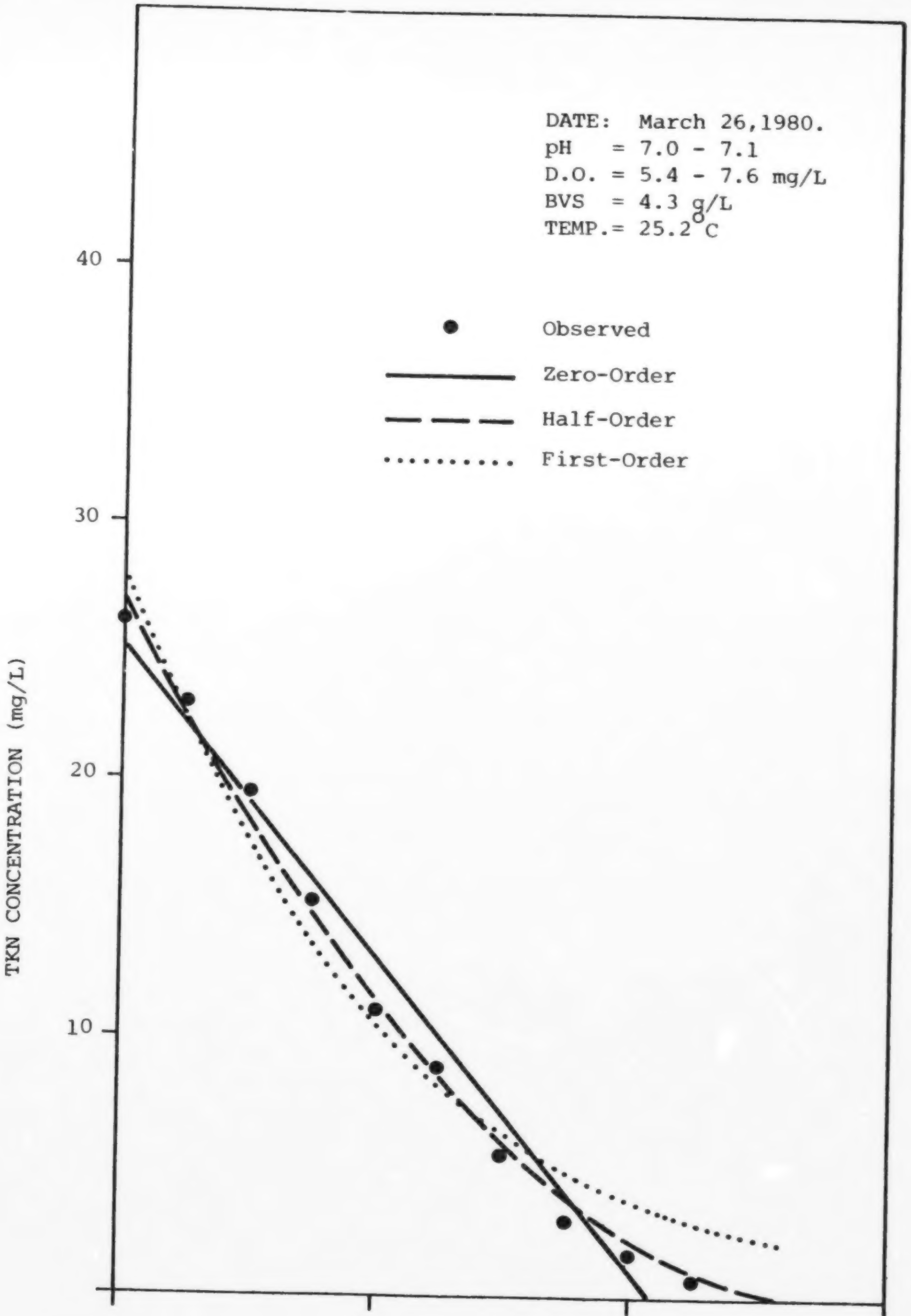

FIGURE 8.12 RESULTS OF FLUIDIZED BED BATCH KINETIC RUN 3

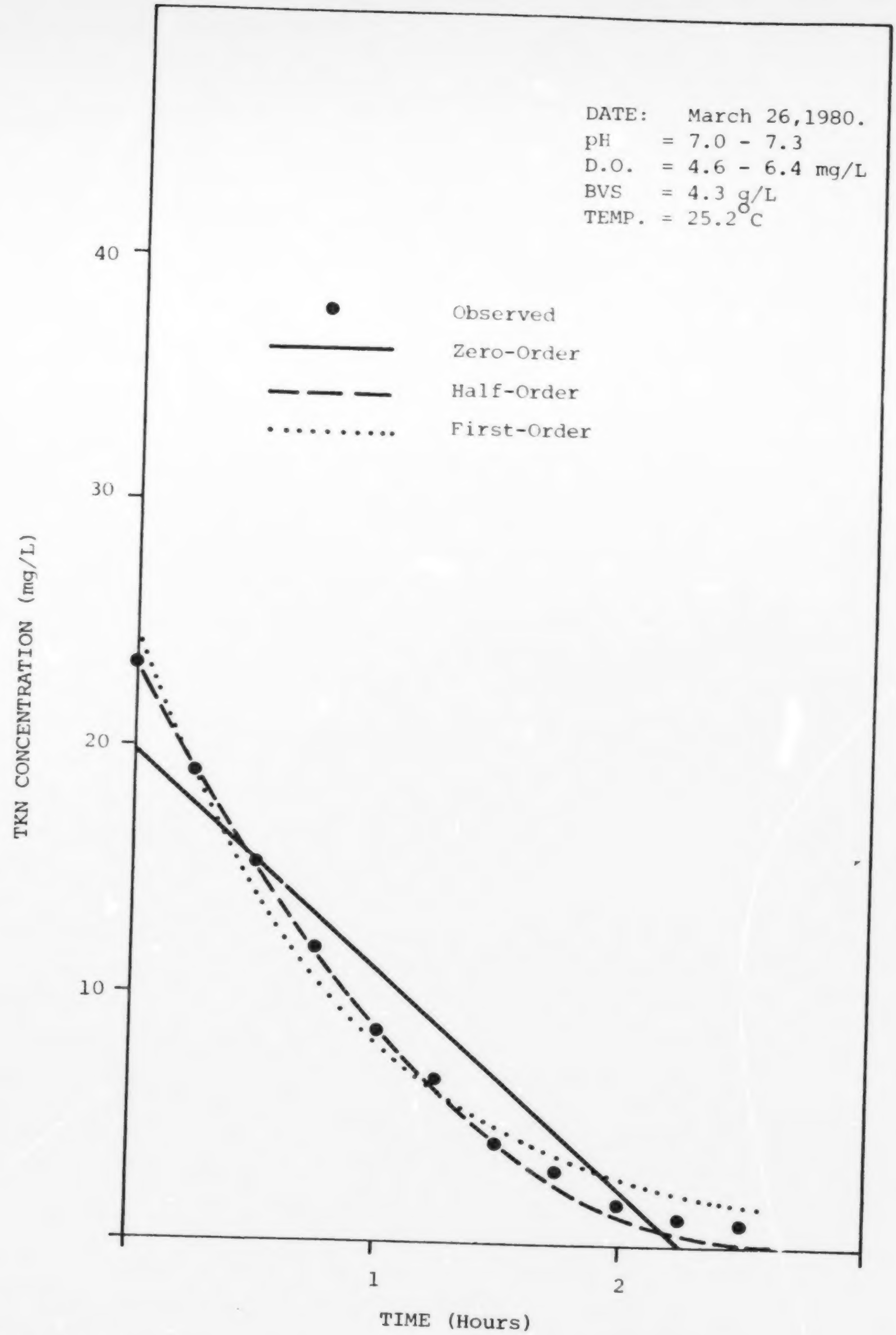

FIGURE 8.13 RESULTS OF FLUIDIZED BED BATCH KINETIC RUN 4

Several investigators(51,52,53,54,55) have examined the effect of diffusion from the bulk liquid to the biofilm in fixed film biological reactors and have concluded that intrinsic zero-order reactions are transformed to bulk half-order reactions in biofilms partially penetrated by substrate. Lamotta(54) demonstrated that the observed reaction order of an n^{th}-order reaction significantly affected by pore diffusion would be:

$$\text{Observed Order} = \frac{(n+1)}{2} \tag{8.2}$$

where n is the intrinsic reaction order.

The intrinsic zero-order nature of the biological ammonia oxidation reaction has been well documented for concentrations of ammonia-nitrogen down to relatively low levels (1 to 3 mg/L)(46,56). The zero-order nature of the nitrification reaction was confirmed by suspending fluidized bed biomass in a mixed, aerated solution of ammonium chloride under constant pH, temperature and dissolved oxygen conditions and monitoring the decline in the ammonia concentration. The results of one such test are shown in Figure 8.14.

Based on the results of the batch kinetic experiments, it is apparent that diffusion limitations are significant in the fluidized bed process under nitrifying conditions despite the thin biofilms ($<100\ \mu m$) present in the biological reactor.

To confirm the half-order nature of the nitrification kinetics, the pseudo-steady state data were fitted to a half-order kinetic model. For the CSTR mixing regime identified in the fluidized bed system under recycle conditions (Section 3.3), the nominal hydraulic retention time (HRT) and the fractional conversion of ammonia-nitrogen ($X = (C_o - C)/C_o$, where C_o is the influent NH_3-N concentration and C is the effluent NH_3-N concentration) are interrelated according to:

$$HRT = \frac{C_o X}{-r} \tag{8.3}$$

where r is the reaction rate.

For a half-order reaction, the rate equation is:

$$-r = -\frac{dC}{dt} = kC^{1/2} \tag{8.4}$$

where k is the reaction rate constant.

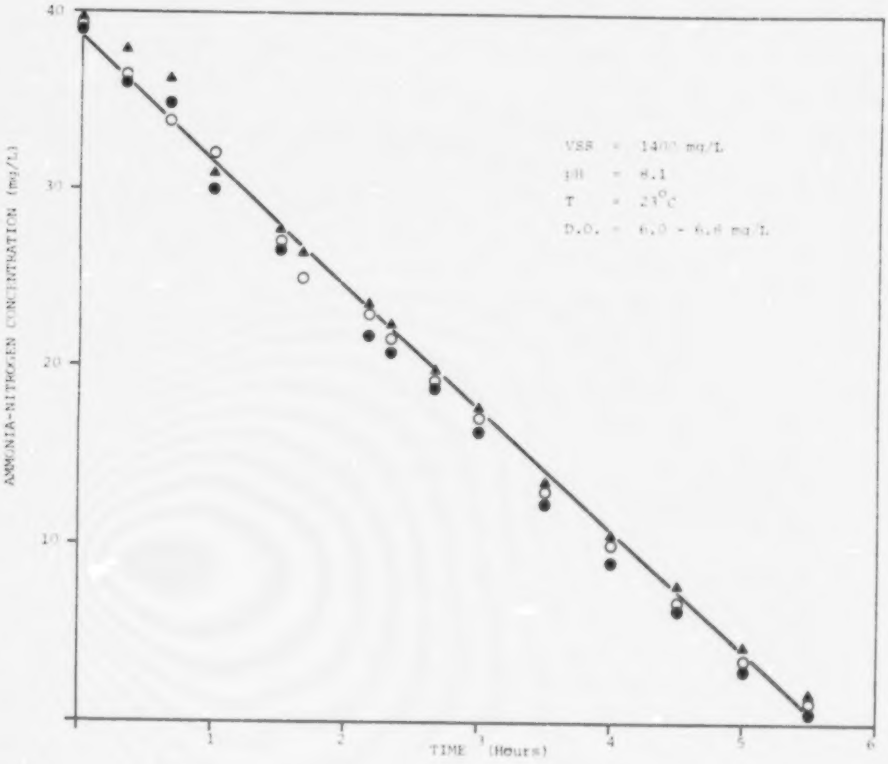

FIGURE 8.14 BATCH NITRIFICATION KINETICS WITH SUSPENDED FLUIDIZED BED BIOMASS

Combining equations (8.3) and (8.4), the relationship for a half-order reaction in a CSTR becomes:

$$\mathrm{HRT} = C_o X/(kC^{1/2}) \qquad (8.5)$$

The pseudo-steady state data are plotted according to the relationship given in equation (8.5) in Figure 8.15.

The data in Figure 8.15 exhibit substantial variation. Much of the variation may be because much of the steady state data was collected at high effluent NH_3-N concentrations. In addition, the lack of true steady state in the process may have influenced these results.

Despite the scatter in the results, the rate constant, k, derived from the pseudo-steady state data in Figure 8.15 agrees with the half-order rate constants based on the batch kinetic runs. The rate constants are compared in Table 8.5. In the case of the rate constant from the batch kinetic experiments, the value of k derived from Figures 8.10 to 8.13 was corrected to eliminate the effect of the sand separation tank (Figure 3.1) on the measured rate of ammonia nitrogen degradation.

TABLE 8.5 HALF-ORDER RATE CONSTANTS FOR BIOLOGICAL NITRIFICATION

Source	Rate Constant, k ($mg^{1/2}\cdot L^{-1/2}\cdot h^{-1}$)
Batch Run 1	9.26
Batch Run 2	7.95
Batch Run 3	13.72
Batch Run 4	12.92
Mean - Batch Runs	10.96
95 percent Confidence Limits Batch Runs	7.65 - 14.66
Pseudo-Steady State Data	9.90

8.5 Effect of Kinetics on Fluidized Bed Nitrification

The concentration dependency resulting from the biofilm diffusion limitations identified during the kinetic modelling of the fluidized bed system is important in the design of a separate stage fluidized bed nitrification system, due to the CSTR mixing regime produced by the oxygenator recycle loop. The CSTR reactor must be larger than a plug flow reactor to provide an equivalent removal efficiency when the reaction rate is concentration dependent(29).

Based on the mean reaction rate constant derived from the batch kinetic experiments ($k = 10.96\ mg^{1/2}\cdot L^{-1/2}\cdot h^{-1}$) which was more precisely defined than that derived from the pseudo-steady state experiments, the effect of influent concentration (C_o) and hydraulic retention time (HRT) on the effluent ammonia-N concentration is

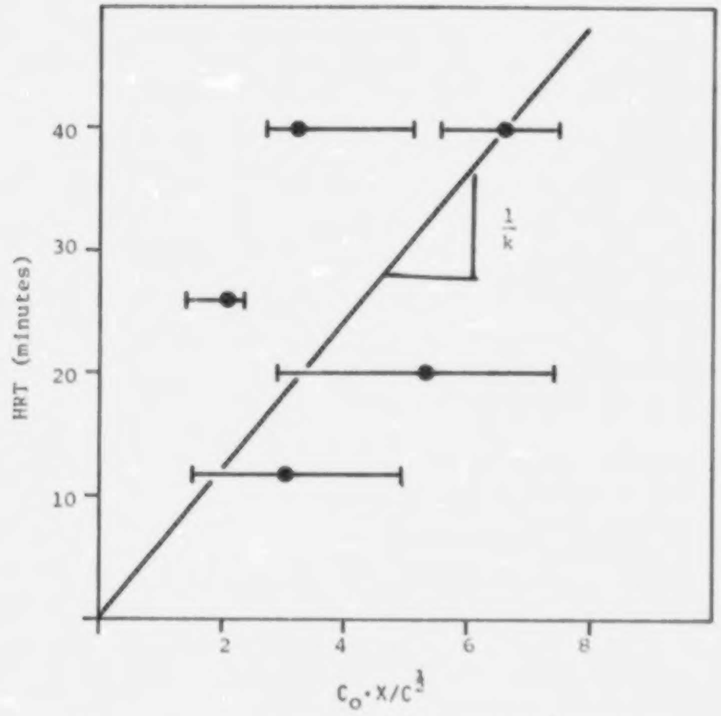

FIGURE 8.15 PSEUDO-STEADY STATE NITRIFICATION RATE DATA

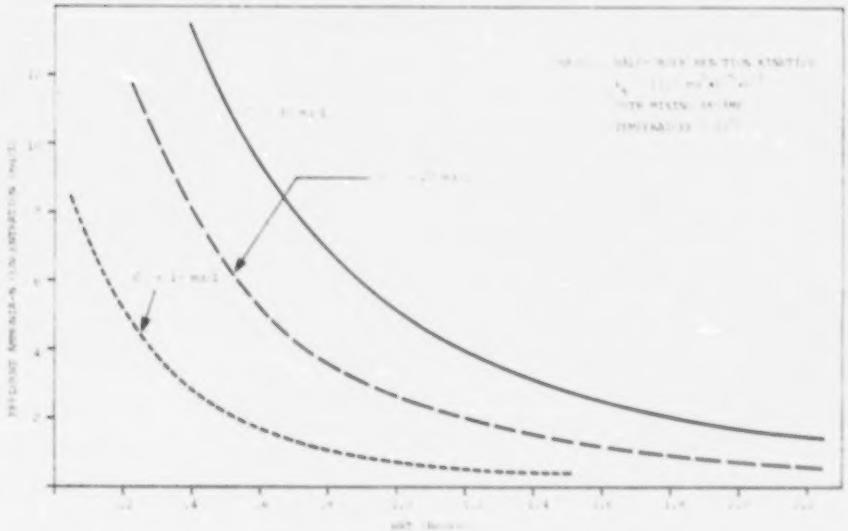

FIGURE 8.16 PREDICTION OF EFFLUENT QUALITY FROM SEPARATE STAGE FLUIDIZED BED NITRIFICATION

shown in Figure 8.16. The predicted effluent quality, based on Figure 8.16, is compared with the actual average effluent quality from the five pseudo-steady state runs in Table 8.6. Although the kinetic model appears to overestimate the effluent concentration at high effluent ammonia-nitrogen concentrations and underestimate the concentration at lower effluent concentrations, the agreement between the predicted and actual effluent quality is relatively good considering the variability of the data. The predicted effluent concentration is within ± 2.5 mg/L over a range of effluent concentrations from 3.1 to 16.0 mg/L NH_3-N.

TABLE 8.6 COMPARISON OF PREDICTED AND ACTUAL EFFLUENT AMMONIA-NITROGEN CONCENTRATIONS

Pseudo-Steady State Run	Effluent Ammonia-Nitrogen Concentration (mg/L)		
	Predicted	Actual	Difference
1	12.2	9.4	+2.6
2	18.0	16.0	+2.0
3	7.0	8.4	-1.4
4	1.1	3.1	-2.0
5	1.8	4.0	-2.2

Based on the half-order kinetic model, the operating conditions required to attain various effluent ammonia-nitrogen concentrations are summarized in Table 8.7. It is evident that a stringent effluent ammonia-N requirement ($\leq$1.0 mg/L) significantly influences fluidized bed process economics in terms of the reactor volume required for a high degree of ammonia-N removal.

It is difficult to compare the efficiency of the fluidized bed process for separate stage nitrification with design data for similar systems. Separate stage suspended growth systems are typically designed to operate at a BOD_5/TKN ratio of 1.5 to 3.0 due to the problems associated with biological solids separation at low influent organic carbon concentrations. The higher influent BOD_5 typically applied to suspended growth systems has a significant effect on the fraction of nitrifiers present and, thus, on the specific removal rates attained.

TABLE 8.7 OPERATING CONDITIONS FOR SEPARATE STAGE FLUIDIZED BED NITRIFICATION PROCESS TO ATTAIN 5 mg/L AND 1 mg/L EFFLUENT AMMONIA-NITROGEN CONCENTRATION (T = 25°C)

Effluent NH_3-N (mg/L)	Influent NH_3-N (mg/L)	HRT* (min)	Unit Loading** (g NH_3-N/g BVS•d)	Volumetric Loading (kg NH_3-N/m^3•d)	Specific Removal** Rate (g NH_3-N/g BVS•d)
5	30	60	0.24	0.72	0.20
	20	37	0.26	0.78	0.19
	10	12	0.40	1.20	0.20
1	30	159	0.09	0.27	0.09
	20	104	0.09	0.28	0.09
	10	49	0.10	0.29	0.09

* HRT based on empty bed reactor volume and secondary effluent feed rate.

** Loading and removal based on BVS = 3.0 g/L.

Similarly, other separate stage fluidized bed nitrification studies(7,57) have been conducted at higher organic carbon to nitrogen ratios. Jeris et al(7) operated at a COD/NH_3-N ratio of 3.7; no BOD_5 data were provided. The unpublished data of Sutton et al(57) were collected at a BOD_5/NH_3-N ratio of approximately 2.0. In both cases significantly higher bed volatile solids concentrations were developed in the fluidized bed (5200 to 8500 mg/L); however, the fraction of nitrifiers would be expected to be proportionally lower. Jeris et al(7) reported a specific removal rate of approximately 0.30 g NH_3-N/g BVS•d at 24°C, approaching this study's maximum rate. Based on the results of these authors, there was no difficulty maintaining an effluent ammonia-N concentration of less than 1 mg/L. The unpublished results of Sutton et al(57) indicate removal rates more in order with these observed in the Burlington pilot plant (0.05 to 0.08 g NH_3-N/g BVS•d at a temperature of 12 to 17°C).

In Table 8.8, an attempt has been made to compare various biological nitrification systems based on the hydraulic retention time required to reduce the influent ammonia-N concentration from 20 mg/L to less than 1 mg/L at 25°C. The fluidized bed nitrification data of Jeris et al(7) and Sutton et al(57) are included for comparison. The reactor volume based on the results of Jeris et al(7) is an order of magnitude less than the volume calculated on the basis of results from the Burlington pilot plant studies. The

TABLE 8.8 COMPARISON OF HYDRAULIC RETENTION TIMES IN SEPARATE STAGE NITRIFICATION SYSTEMS
Basis: C_o = 20 mg/L NH_3 - N; C $\leq$1 mg/L NH_3 - N; T = 25°C

Process	Source	Carbon-Nitrogen Ratio	Hydraulic Retention Time (minutes)
Fluidized Bed	This Study	$BOD_5/TKN < 0.4$	104
	Jeris *et al*(7)	COD/TKN = 3.7	11(a)
	Sutton *et al*(57)	BOD_5/NH_3-N = 2	54(b)
Packed Bed Reactor	Gasser *et al*(58)	BOD_5/NH_3-N = 0.4	90(c)
RBC	Antonie(59)	BOD_5/NH_3-N < 1	90(d) - 280(e)
Complete Mix Activated Sludge	USEPA(46)	BOD_5/TKN = 1.3	180(f)

Assumptions:

(a) Nitrification rate = 0.3 g/g•d; BVS = 8.5 g/L.

(b) Nitrification rate at 25°C = 0.085 g NH_3-N/g•d based on average of measured rate at 13°C (0.05 g/g•d) x 1.40 and at 17°C (0.08) x 1.23 (temperature correction based on Figure 8.17(61)). Average BVS = 6 g/L.

(c) Nitrification rate = 0.3 kg/m^3•d at 19 to 28°C. Support medium: 5 mm gravel.

(d) Hydraulic loading = 82 L/m^2•d to achieve 95% NH_3-N removal. T >13°C. Specific Surface = 200 m^2/m^3 in a 4-stage RBC.

(e) Nitrification Rate = 0.485 kg/1000m^2•d to achieve 1 mg/L effluent NH_3-N removal. T >13°C. Specific Surface = 200 m^2/m^3 in a 4-stage RBC.

(f) Peak nitrification rate = 0.3 g/g•d at 25°C. Dissolved oxygen concentration 2 mg/L. Aeration tank MLVSS = 2000 mg/L.

preliminary results of Sutton *et al*(57) agree more closely with the Burlington results. In the calculation of the HRT required based on the results of Sutton *et al*(57), it was assumed that there would be a significant increase in the nitrification rate at a temperature of 25°C. This would have to be confirmed.

Comparison of the preliminary results of Sutton *et al*(57) and the results from the Burlington studies indicate that there may be an advantage to operating the biological nitrification process at a higher BOD_5/N ratio. Huang(60) has also noted an improvement in the nitrification rate in an attached growth biological treatment system in the presence of a small amount of biodegradable organic carbon ($BOD_5 \approx 10$ mg/L).

Comparison of the HRT required for other biological nitrification processes (packed bed reactor, RBC and activated sludge) with the HRT calculated from the Burlington results at the low BOD_5/TKN ratio do not reflect the large space advantages of the fluidized bed process claimed by Jeris et al(7) at the high ratio. However, it should be noted that a number of assumptions were made in calculating the HRT required for the alternative systems. For example, in the packed bed reactor, it was assumed that the nitrification rate was not significantly reduced at low effluent NH_3-N concentrations ($\leq$1.0 mg/L). The nitrification rate used in calculating the required HRT was based on data collected at effluent NH_3-N concentrations of 1.3 to 2.0 mg/L(58). A direct comparison of these nitrification systems, based on parallel process operation, would be necessary to verify the data presented in Table 8.8.

8.6 Other Factors Affecting Nitrification

8.6.1 Alkalinity Consumption. Alkalinity is consumed during the biological oxidation of ammonia. The theoretical requirement is 7.1 mg alkalinity (as $CaCO_3$) per mg NH_3 -N oxidized(46). In experimental assessments of attached growth nitrification systems, the alkalinity consumption ratio has varied from 6.3 to 7.4 mg $CaCO_3$/mg NH_3 -N(46).

Based on the pseudo-steady state operating data, the average alkalinity consumption ratio was 7.99 ± 0.38 mg $CaCO_3$/mg NH_3-N. This is a slightly higher ratio than that typically cited for nitrification processes. The alkalinity measurements were conducted on grab samples of influent and effluent from the fluidized bed; the ammonia-N removal data were based on 24-hour composite samples of influent and effluent. The difference in sampling technique, combined with experimental error, were probably responsible for the discrepancy between the theoretical and measured alkalinity consumption ratio.

8.6.2 Oxygen Consumption. The theoretical oxygen requirement for nitrification, neglecting synthesis, is 4.57 mg O_2/mg NH_4^+-N oxidized(46). The theoretical oxygen requirement for the combined synthesis and oxidation is 4.19 mg O_2/mg NH_3-N(46). The typical design oxygen consumption ratio is 4.6 mg O_2/mg NH_3-N(46).

Based on 12 daily records of influent and effluent dissolved oxygen concentration given in Table 8.9, the oxygen consumption during the pseudo-steady state nitrifi-

TABLE 8.9 OXYGEN UTILIZATION FOR NITRIFICATION

Date	Reactor Feed Rate (L/min)	Recycle Rate (L/min)	Dissolved Oxygen (mg/L)		Ammonia-N (mg/L)		Oxygen Utilized (mg O_2/mg NH_3-N)
			Influent	Effluent	Influent	Effluent	
Jan. 9/80	18.16	45.4	10.4	4.8	21.5	16.1	3.63
Jan. 15/80	18.16	45.4	11.2	5.0	23.8	18.1	3.81
Jan. 23/80	10.44	53.1	10.0	5.0	33.0	24.9	3.75
Feb. 6/80	10.44	53.1	14.4	5.0	21.9	9.5	4.61
Feb. 13/80	10.44	53.1	14.8	5.0	21.0	5.6	3.87
Feb. 20/80	10.44	53.1	15.6	5.0	24.5	6.1	3.51
Feb. 27/80	18.16	45.4	17.6	5.0	21.7	10.6	3.97
March 5/80	18.16	45.4	12.0	5.0	20.9	15.9	4.90
March 12/80	18.16	45.4	14.4	5.0	20.9	11.4	3.46
March 19/80	18.16	45.4	17.6	5.0	25.1	12.5	3.50
April 2/80	5.31	58.2	10.0	5.2	26.6	8.9	3.24
April 16/80	5.31	58.2	6.7	5.0	8.2	3.2	4.07
						Mean =	3.86

cation runs was 3.86 ± 0.31 mg O_2/mg NH_3-N oxidized. There was no significant oxygen demand related to the degradation of organic compounds.

The measured oxygen consumption is slightly lower than values typically cited for nitrifying biological systems. The discrepancy between the measured and theoretical values is probably related to inaccuracies in the continuous measurement of influent and effluent dissolved oxygen concentrations.

8.6.3 Solids Yield and Process SRT. As the fluidized bed influent contained no significant quantities of biodegradable organic carbon, the process sludge yield was relatively low and the system was operated at a relatively long SRT. It was unnecessary to waste biomass from the system to control the higher growth rate associated with heterotrophic organisms. The SRT during the pseudo-steady state runs ranged from 25 to 39 days. This does not represent the steady state system SRT as the biomass concentration in the reactor did not attain equilibrium.

Based on the process SRT and the system ammonia removal efficiency, the solid yield ranged from 0.08 to 0.13 g VS/g NH_3-N. The calculated solid yield agrees with the estimated theoretical yield of nitrifying organisms, 0.15 g/g NH_3-N oxidized(46).

8.6.4 Effect of Temperature. All of the pseudo-steady state runs at Burlington were carried out at 25°C, so no data are available concerning the effect of temperature on the fluidized bed nitrification process.

There is general agreement in the literature that attached growth nitrification systems are less temperature-sensitive than similar suspended growth processes. The effect of temperature on attached growth and suspended growth systems, based on the work of a number of investigators, is shown in Figure 8.17(61). However, the precise impact of temperature on an attached growth nitrification system is less well defined.

Investigations at Rutgers University(67) found that the nitrification capacity of an RBC was unaffected by temperatures between 15 to 26°C. Similarly, Antonie(59) found no temperature dependency in an RBC above 13°C, although the rates were significantly reduced as the temperature was reduced to 4°C. Operating data collected with packed bed nitrification systems indicate that temperature has a significant impact on process performance over a range from 15 to 27°C(58,68,69). The hydraulic retention time in the packed bed reactor must be increased from about 55 minutes at 27°C to 200 minutes at 15°C to maintain the same efficiency.

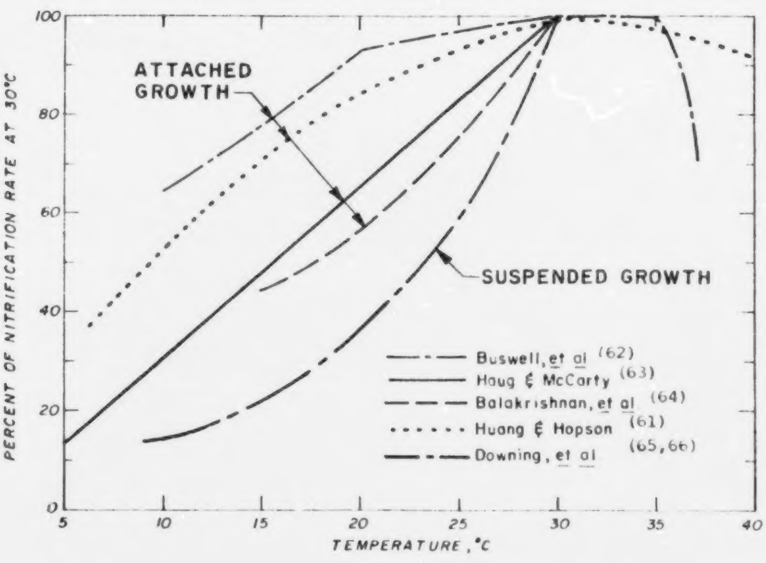

FIGURE 8.17 COMPARISON OF EFFECT OF TEMPERATURE ON NITRIFICATION IN SUSPENDED GROWTH AND ATTACHED GROWTH SYSTEMS

From USEPA(46)

There are insufficient data to determine the effect of temperature on the fluidized bed nitrification process. Further investigations should be carried out to define the effect of temperature on process design.

8.7 Summary

Although the separate stage fluidized bed nitrification system appears to offer some advantages in terms of reactor size relative to other suspended growth and fixed film nitrification processes, the results collected at Burlington using high quality secondary effluent (BOD_5/TKN <0.4) did not substantiate the order of magnitude reduction in reactor size suggested by Jeris et al(7). The high biomass concentrations reported by Jeris et al(7) (8500 mg/L) and Sutton et al(57) (5200 to 7000 mg/L) could not be attained at Burlington with the low BOD_5 secondary effluent used as feed. Therefore, despite specific nitrification rates similar to those reported by Sutton et al(57), significantly larger reactors would be necessary to attain the same degree of treatment. It is possible that the use of smaller support media, with greater bed expansion and the presence of a small amount of biodegradable organic carbon, would result in increased reactor biomass concentrations without significantly affecting the specific nitrification rate.

A lengthy start-up time was necessary before an active nitrifying population was established in the fluidized bed reactor and a steady state condition in terms of population dynamics was reached. Under pseudo-steady state conditions, alkalinity consumption, oxygen consumption and solids yield in the fluidized bed were similar to published values for other biological nitrification processes.

Significant diffusional limitations within the biofilms were identified. The half-order ammonia-nitrogen concentration dependency, in conjunction with the CSTR mixing regime present in the fluidized bed system under recycle conditions ($Q_R Q_I \geq 2$), strongly influences process design. Although 75 percent ammonia-nitrogen removal from an influent concentration of 20 mg/L NH_3-N could be attained at a reactor HRT of only 37 minutes, biological treatment to attain a high quality effluent (≥ 1 mg/L NH_3-N) required an increase in the reactor HRT to more than 100 minutes. Improved oxygenator design to allow operation at low recycle ratios, and reactor design to ensure a plug flow mixing regime in the reaction vessel, would help minimize the effect of biofilm diffusion limitations and maximize the rate of nitrification attained.

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ABBREVIATIONS AND SYMBOLS

a_t - white noise sequence
Alk - alkalinity
ATP - adenosine triphosphate
Avg - average
b - system delay period
B - backward shift operator (e.g. $B^n Y_t = Y_{t-n}$)
BOD_5 - biochemical oxygen demand
BVS - bed volatile solids
C - concentration
C_o - influent concentration
CMHC - Canada Mortgage and Housing Corporation
COD - chemical oxygen demand
CSTR - continuous stirred tank reactor
d_{10} - effective size (diameter)
d_{60}/d_{10} - uniformity coefficient
D_p - particle diameter
D_L - diffusivity coefficient
$D/\mu L$ - dispersion number
DO - dissolved oxygen concentration
$\frac{dC}{dt}$ - reaction rate (rate of charge of concentration)
E - removal efficiency
E/E_o - external residence time distribution
E.A. - extended aeration
f - fraction of nitrifiers
F - filterable
F/M - food to mass ratio
$FeCl_3$ - ferric chloride
FOC - filterable organic carbon
FOC_E - effluent filterable organic carbon
FOC_I - influent filterable organic carbon
FOC_L - filterable organic carbon mass loading

g - process gain
g/L - gram per litre
g/g BVS•d - process loading
g O_2/g BOD_5 - specific oxygen utilization rate
HRT - hydraulic retention time
k - reaction rate constant
ln - natural logarithm
MLVSS - mixed liquor volatile suspended solids
n - reaction order
N - number of CSTR's
NH_3-N - ammonia nitrogen (also NH_4-N, NH_4^+-N)
NO_2-N - nitrite nitrogen
NO_3-N - nitrate nitrogen
p, d, f - noise model orders
Q - flow rate
Q_o - primary effluent flow rate
Q_D - dilution water flow rate
Q_I - reactor feed rate ($Q_o + Q_D$)
Q_R - reactor recycle rate
Q_T - total reactor flow rate ($Q_I + Q_R$)
R - recycle ratio (Q_R/Q_I)
r - reaction rate (dC/dt)
RBC - rotating biological contactor
SRT - solids retention time
SRT_a - actual solids retention time
SRT_c - conventional solids retention time
SRT_m - modified solids retention time
SS - suspended solids
T - temperature
TKN - total kjeldahl nitrogen
TSS - total suspended solids
USEPA - United States Environmental Protection Agency
v - velocity

VS	-	volatile solids
VSS	-	volatile suspended solids
WPCP	-	Water Pollution Control Plant
WRC	-	Water Research Centre
WTC	-	Wastewater Technology Centre
X	-	fractional conversion
Y	-	solids yield
Y_a	-	actual solids yield
Y_c	-	conventional solids yield
Y_m	-	modified solids yield
τ	-	time constant